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# ANALYTICAL STUDY OF CATALYTIC REACTORS FOR HYDRAZINE DECOMPOSITION

by

ARTHUR S. KESTEN

prepared for

NATIONAL AERONAUTICS AND SPACE ADMINISTRATION

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RESEARCH LABORATORIES
United Aircraft Corporation
East Hartford, Connecticut

# Analytical Study of Catalytic Reactors for Hydrazine Decomposition

First Annual Progress Report

April 15, 1966 - April 14, 1967

Contract No. NAS 7-458

TABLE OF CONTENTS


| | Page |
|---|---|
| ABSTRACT | i |
| FOREWORD | ii |
| SUMMARY | 1 |
| INTRODUCTION | 2 |
| DISCUSSION | 3 |
| Steady-State Model | 4 |
| Transient Model | 8 |
| Kinetics Information | 13 |
| Results of Calculations | 15 |
| Steady-State Results | 15 |
| Transient Results | 16 |
| REFERENCES | 19 |
| LIST OF SYMBOLS | 21 |
| APPENDIX I | 24 |
| APPENDIX II | 27 |
| FIGURES | 34-83 |

## ABSTRACT

An analytical study of catalyzed hydrazine decomposition reaction chambers was performed in order to establish procedures capable of predicting the steady-state and transient behavior of the system. The study included the development of computer programs to calculate the temperature and reactant concentration distribution as functions of time and axial position in typical reaction chamber configurations. The computer programs evolved are based upon models of the reactor system which consider both thermal and catalytic decomposition of reactants, along with simultaneous heat and mass transfer between the free-gas phase and the gas within the pores of the catalyst pellets. Results obtained using these computer programs are in excellent agreement with data available from small scale engine firings using Shell 405 catalyst thus permitting not only correlation of the test data but also indirect determination of hydrazine and ammonia decomposition kinetics on 405 catalyst.

## FOREWORD

This work was performed by United Aircraft Research Laboratories for the National Aeronautics and Space Administration under Contract NAS 7-458 initiated April 15, 1966.

Included among those who cooperated in performance of the work under Contract NAS 7-458 were Dr. A. S. Kesten, Program Manager, Dr. W. G. Burwell, Chief, Kinetics and Thermal Sciences Section, Mr. D. B. Smith, and Mrs. E. Smith of UARL.

This work was conducted under program management of the NASA Chief, Liquid Propulsion Experimental Engineering Systems, NASA Headquarters, Washington, D. C., and the Technical Manager was Mr. T. W. Price, Jet Propulsion Laboratory, Pasadena, California

Report F910461-12

# Analytical Study of Catalytic Reactors for Hydrazine Decomposition

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## SUMMARY

The Research Laboratories of United Aircraft Corporation under Contract NAS 7-458 with the National Aeronautics and Space Administration are performing an analytical study of catalytic reactors for hydrazine decomposition. This first annual technical report summarizes work performed under this continuing contract from April 15, 1966 to April 14, 1967. Work during this period has included the development of computer programs which are used to calculate the temperature and concentrations of hydrazine and ammonia as functions of time and axial position in typical reaction chamber configurations. The computer programs are based upon steady-state and transient models of the reactor system which consider both thermal and catalytic decomposition of reactants, along with simultaneous heat and mass transfer between the free-gas phase and the gas within the pores of the catalyst pellets. The steady-state model contains a detailed mathematical description of the processes of heat and mass transfer in the pores of the catalyst pellets, while the transient model uses over-all coefficients to evaluate the effects of heat and mass diffusion on reaction rates in the porous catalyst particles. These coefficients are defined in terms of the physical properties and operating parameters of the reactor system.

The computer programs representing the steady-state and transient models of the reactor system have been used to calculate temperature and reactant concentration distributions for various initial bed temperatures, feed temperatures, chamber pressures, mass flow rates, injector locations, and catalyst size distributions. The results have been compared, wherever possible, to available measurements of temperature profiles and exhaust concentrations made during firings of typical hydrazine engine reaction chambers employing Shell 405 catalyst. Excellent agreement between theoretical and experimental results is indicated.

This agreement has permitted verification of kinetics information for use in both the steady-state and transient programs. In particular, estimates have been permitted of the rate laws and associated rate constants governing the heterogeneous, vapor-phase decomposition of hydrazine and ammonia on the Shell 405 catalyst.

# INTRODUCTION

Effective design of distributed-feed catalyzed monopropellant hydrazine rocket engines and gas generators requires accurate procedures for predicting the effects of the design parameters of the reactor system on the steady-state and transient performance of the system. Prior to the present undertaking, this general capability did not exist, although simplified methods had been developed for predicting the steady-state performance of idealized catalytic reactors (Refs. 1 and 2). These simplified methods do not adequately describe the combined processes of heat transfer, diffusion, and chemical reaction and are restricted to consideration of a single decomposition reaction at any axial location within the reaction chamber. To overcome these deficiencies, a more comprehensive theoretical analysis was deemed necessary. Preliminary investigations at United Aircraft Research Laboratories demonstrated the feasibility of such an analysis and, further, indicated that both transient and steady-state performance characteristics could be predicted.

Based upon the above investigations, a comprehensive analytical program was formulated with the objectives of (a) developing computer programs for predicting the temperature and concentration distributions in monopropellant hydrazine catalytic reactors, and (b) performing calculations using these computer programs to demonstrate the effects of various system parameters on the performance of the reactor and, ultimately, to establish guidelines for reactor design. During the first year of the contract period attention has been focused on describing the steady-state and transient behavior of a reactor system in which complete radial mixing in the free-gas (or liquid) phase was assumed. Currently, the steady-state program is being extended to include radial as well as axial variations in temperature and concentrations. This portion of the program, scheduled for completion in April of 1968, will permit an analysis of various injection schemes which exhibit radial nonuniformities. Effort during the first annual reporting period is described in detail in succeeding sections of this report.

## DISCUSSION

The analysis of a hydrazine engine reaction system carried out to date pertains to a reaction chamber of arbitrary cross section packed with catalyst particles into which liquid hydrazine is injected at arbitrarily selected axial locations. At these locations, hydrazine injection is taken as uniform across the cross section of the chamber. Catalyst particles are represented as "equivalent" spheres with a diameter taken as a function of the particle size and shape. Both thermal and catalytic vapor phase decomposition of hydrazine and ammonia are considered in developing equations describing the concentration distributions of these reactants. Diffusion of reactants from the free-gas phase to the outside surface of the catalyst pellets is taken into account. Since the catalyst material is impregnated on the interior and exterior surfaces of porous particles, the diffusion of reactants into the porous structure must also be considered. In addition, the conduction of heat within the porous particles must be taken into account since the decomposition reactions are accompanied by the evolution or absorption of heat.

A treatment of the transport processes described above constitutes a general model of the reaction chamber. The analysis of the steady-state performance of a reactor based on this model can be carried out in a straight-forward manner. However, the analysis of the transient behavior of the system using this general model is quite complicated since both the reactant concentrations and the rate constants for decomposition, which are exponentially dependent on temperature, will be functions not only of time, but also of position within the catalyst particles.

To circumvent these complications, the transient behavior of the system can be analyzed by employing a second model which takes heat and mass diffusion within the catalyst particles into account by considering their effects on the rates of the catalytic reactions through the "utilization factor." This factor, which is defined as the ratio of actual reaction rate to that which would be produced by the catalyst particle if negligible temperature and concentration gradients existed within the particle, can be calculated as a function of temperature and reactant concentration at the outside surface of a catalyst particle at steady-state. If it is assumed that, even under transient conditions, utilization factors are functions only of surface temperatures and concentrations, the analysis of the transient behavior of the reactor system becomes tractable. Using this method then, the temperature and reactant concentrations can be described as functions of time and position in the reaction chamber.

Included in succeeding sections of this report are detailed descriptions of (a) the development of the computer programs representing the steady-state and transient models of the reactor system, (b) the establishment of kinetics information pertinent to Shell 405 catalyst particles (Refs. 3, 4, 5, and 6) for use in both of these programs, and (c) the use of these programs to calculate temperature and reactant concentration distributions for various reactor operating conditions.

Steady-State Model

In developing the present steady-state model, the temperature and reactant concentrations in the interstitial phase (i.e., the free-fluid phase as distinguished from the gas phase within the porous particles) are assumed to vary only with axial distance along the bed. In the entrance region of the reaction chamber, where the temperature is low enough to permit the existence of liquid hydrazine, vaporization of liquid is assumed to occur as a result of decomposition of vapor hydrazine within the pores of the catalyst particles. That is, catalytic reaction is assumed to be fast enough to keep liquid hydrazine from wetting the pores of the particles; the hydrazine concentration at the surface of the catalyst particles at any axial location in the entrance region is then computed from the vapor pressure of liquid hydrazine in the interstitial phase at the same axial location. Neglecting axial diffusion of heat or mass, the change in enthalpy of the interstitial phase in the region where liquid hydrazine is present (i.e., where $h_i < h_i^V$) is related to the concentration gradient at the surface of the porous catalyst particles by

$$G \frac{dh_i}{dz} + H^{N_2H_4} D_p A_p \left(\frac{dc_p^{N_2H_4}}{dx}\right)_s + F(h_i - h_F) = 0 \qquad \text{for } h_i \le h_i^V \quad (1)$$

The variation of mass flow rate, G, with axial distance is easily computed from the rate of feed of liquid hydrazine from the distributed injectors into the system. In the region where liquid hydrazine exists at temperatures below the vaporization temperature, the temperature may be obtained from

$$T_i = T_F + \frac{h_i - h_F}{C_f} \qquad \text{for } h_i < h_i^L \quad (2)$$

In the two-phase region, where $T_i = T_{vap}$, the weight-fraction of vapor may be computed from

$$\text{WEIGHT - FRACTION VAPOR} = \frac{h_i - h_i^L}{h_i^V - h_i^L} \qquad \text{for } h_i^L \le h_i \le h_i^V \quad (3)$$

At the axial position at which the enthalpy of the interstitial phase is just equal to the enthalpy of vapor hydrazine at the boiling point ($h_i = h_i^V$), the fraction of hydrazine injected upstream of that point which has been decomposed is easily calculated from an overall heat balance. The associated amounts of ammonia, nitrogen, and hydrazine formed from decomposition of hydrazine can then be calculated taking the decomposition reaction as

$$2N_2H_4 \longrightarrow 2NH_3 + N_2 + H_2$$

It should be noted that this is the overall reaction scheme determined experimentally for both homogeneous decomposition of hydrazine (Refs. 7, 8, 9) and low pressure heterogeneous decomposition of hydrazine on platinum surfaces (Ref. 10).*

In the remainder of the reaction chamber, where $h_i > h_i^V$, heat is being supplied to the system by homogeneous as well as heterogeneous decomposition of hydrazine. In addition, at sufficiently high temperature, heat is removed from the system by the endothermic decomposition of ammonia. For $h_i > h_i^V$ then, the change in enthalpy with axial distance is related to the concentration gradients at the surface of the porous catalyst particles by

$$G\frac{dh_i}{dz} + H^{N_2H_4} r_{hom}\,\gamma + A_p\left[H D_p\left(\frac{dc_p}{dx}\right)_s\right]^{N_2H_4} + A_p\left[H D_p\left(\frac{dc_p}{dx}\right)_s\right]^{NH_3} + F(h_i - h_F) = 0 \qquad (4)$$

The changes in reactant weight fractions in the interstitial phase with axial distance are related to the concentration gradients at the surface of the porous catalyst particles by

$$\frac{dw_i^{N_2H_4}}{dz} = \frac{1}{G}\left\{F - r_{hom}\delta - A_p\left[D_p\left(\frac{dc_p}{dx}\right)_s\right]^{N_2H_4}\right\} - \frac{F c_i^{N_2H_4}}{G\rho_i} \qquad (5)$$

$$\frac{dw_i^{NH_3}}{dz} = \frac{1}{G}\left\{r_{hom}\,\delta\,\frac{M^{NH_3}}{M^{N_2H_4}} + A_p\left[D_p\left(\frac{dc_p}{dx}\right)_s\right]^{N_2H_4}\frac{M^{NH_3}}{M^{N_2H_4}} - A_p\left[D_p\left(\frac{dc_p}{dx}\right)_s\right]^{NH_3}\right\} - \frac{F c_i^{NH_3}}{G\rho_i} \qquad (6)$$

*It is more commonly assumed, without benefit of experimental evidence, that the decomposition reaction is $3\,N_2H_4 \longrightarrow 4\,NH_3 + N_2$, followed by dissociation of one of the four ammonia molecules to nitrogen and hydrogen. This two-step process leads to the same overall reaction cited above but assumes that a minimum of 25 percent of the ammonia produced by hydrazine decomposition also decomposes. The fractional ammonia dissociation, f, calculated assuming the validity of the two-step process is related to the fractional ammonia dissociation calculated in the present report by

$$(f)_{\text{two-step process}} = \frac{3\,(f)_{\text{present report}} + 1}{4}$$

$$\frac{dw_i^{N_2}}{dz} = \frac{1}{G}\left\{\frac{1}{2} r_{hom}\delta \frac{M^{N_2}}{M^{N_2H_4}} + \frac{A_p}{2}\left[D_p\left(\frac{dc_p}{dx}\right)_s\right]^{N_2H_4}\frac{M^{N_2}}{M^{N_2H_4}} + \frac{A_p}{2}\left[D_p\left(\frac{dc_p}{dx}\right)_s\right]^{NH_3}\frac{M^{N_2}}{M^{NH_3}}\right\} - \frac{Fc_i^{N_2}}{G\rho_i} \tag{7}$$

$$\frac{dw_i^{H_2}}{dz} = \frac{1}{G}\left\{\frac{1}{2} r_{hom}\delta \frac{M^{H_2}}{M^{N_2H_4}} + \frac{A_p}{2}\left[D_p\left(\frac{dc_p}{dx}\right)_s\right]^{N_2H_4}\frac{M^{H_2}}{M^{N_2H_4}} + \frac{3A_p}{2}\left[D_p\left(\frac{dc_p}{dx}\right)_s\right]^{NH_3}\frac{M^{H_2}}{M^{NH_3}}\right\} - \frac{Fc_i^{H_2}}{G\rho_i} \tag{8}$$

The changes in reactant concentrations with axial distance are then given by

$$\frac{dc_i^J}{dz} = \rho_i \frac{dw_i^J}{dz} + \frac{c_i^J}{\rho_i}\frac{d\rho_i}{dz} \tag{9}$$

where

$$\frac{1}{\rho_i}\frac{d\rho_i}{dz} = \frac{1}{\bar{M}}\frac{d\bar{M}}{dz} - \frac{1}{T_i}\frac{dT_i}{dz} \tag{10}$$

and

$$\frac{1}{\bar{M}}\frac{d\bar{M}}{dz} = -\frac{1}{\sum_J \frac{w_i^J}{M^J}}\sum_J \frac{1}{M^J}\frac{dw_i^J}{dz} \tag{11}$$

The temperature of the interstitial phase in this region is related to the enthalpy by

$$h_i - h_i^V = \int_{T_{vap}}^{T_i} C_f \, dT_i \tag{12}$$

where $C_f$ is the specific heat of the gas mixture and is a function of temperature as well as the concentration of the constituents of the mixture.

The reactant concentration profile in the porous particles at any axial location must satisfy the diffusion equation for mass transport as well as an analogous equation for heat conduction. Neglecting the effect of the translational motion

of the gas stream on diffusion within the porous particles and assuming constant diffusion coefficients, $D_p$, and thermal conductivities, $K_p$, these equations may be written as

$$D_p \nabla^2 c_p - r_{het} = 0 \tag{13}$$

$$K_p \nabla^2 T_p - H r_{het} = 0 \tag{14}$$

Using Eqs. (13) and (14), Prater (Ref. 11) has pointed out that temperature and concentration are related quite simply by

$$T_p - (T_p)_s = - \frac{H D_p}{K_p} \left[ (c_p)_s - c_p \right] \tag{15}$$

The use of this relationship enables the reaction rate, $r_{het}$, to be written as a function of concentration alone instead of concentration and temperature. Equation (13) can then be solved for the concentration at any point in the porous particle in terms of the concentration at the surface of the particle, $(c_p)_s$. The solution is derived in Appendix A as an implicit integral equation given by

$$c_p(x) = (c_p)_s - \left[ \frac{1}{x} - \frac{1}{a} \right] \int_0^x \xi^2 \frac{r_{het}(c_p)}{D_p} \, d\xi - \int_x^a \left[ \frac{1}{\xi} - \frac{1}{a} \right] \xi^2 \frac{r_{het}(c_p)}{D_p} \, d\xi \tag{16}$$

The rate of chemical reaction on the catalyst surfaces, $r_{het}(c_p)$, is given, in general, by (Ref. 12)

$$r_{het}(c_p) = k_o (c_p)_s^{1-n} (c_p)^n \exp \left\{ \gamma\beta \left( 1 - c_p/(c_p)_s \right) / \left[ 1 + \beta \left( 1 - c_p/(c_p)_s \right) \right] \right\} \tag{17}$$

where the parameters, $k_o$, $n$, $\gamma$, and $\beta$ are defined in the list of symbols.

If the reactant concentration at the surface of the particle, $(c_p)_s$, is equal to the concentration in the interstitial phase, $c_i$, Eqs. (16) and (17) can be solved simultaneously with the equations describing the changes in enthalpy and reactant concentrations in the interstitial phase with axial distance to yield the

steady-state temperature and concentration profiles in the reaction chamber. In the liquid region of the chamber, for example, it is assumed that liquid hydrazine wets the outside surface of the catalyst particles so that $(c_p)_s^{N_2H_4} = c_i^{N_2H_4}$ where $c_i^{N_2H_4}$ is the vapor concentration in equilibrium with liquid hydrazine at temperature $T_i$. In the vapor region of the reactor it can be shown that $(c_p)_s^{NH_3} \sim c_i^{NH_3}$ at the temperatures existing in the chamber since the rate of transport of material from the bulk fluid to the outside surface of the catalyst particles is much greater than the rate of dissociation of ammonia within the particles. However, in the case of hydrazine decomposition in the vapor region the rate of catalytic decomposition is so high that $(c_p)_s^{N_2H_4} \ll c_i^{N_2H_4}$ and the reaction rate is governed by the rate of mass transfer from the bulk vapor to the outside surface of the catalyst particles. This rate is given approximately by $(k_c c_i)^{N_2H_4}$ where the mass transfer coefficient, $k_c^{N_2H_4}$, may be estimated from (Ref. 13)

$$k_c^{N_2H_4} = \frac{0.61\,G}{\rho_i}\left(\frac{\mu}{\rho_i D_i^{N_2H_4}}\right)^{-0.667}\left(\frac{G}{A_p\,\mu}\right)^{-0.41} \qquad (18)$$

For hydrazine decomposition in the vapor region, then, the terms $\left[D_p\,(\partial c_p/\partial x)_s\right]^{N_2H_4}$ in Eqs. (4), (5), (6), (7) and (8) can be replaced by $(k_c\,c_i)^{N_2H_4}$.

In the liquid-vapor region the situation is somewhat more complicated since it is difficult to predict whether liquid or a combination of liquid and vapor wets the outside surface of the catalyst particles. Both of these options are presently in the computer program representing the steady-state model. In the case in which both the liquid and vapor are taken to wet the particle surface, it is assumed that, at a given axial location, the fraction of the surface covered by vapor is equal to the weight-fraction of vapor present. Decomposition rates, computed assuming pure liquid surface coverage and then pure vapor coverage, are weighted accordingly. Fortuitously, for the system considered here, the liquid-vapor region is so narrow that the choice of either of these options has negligible effect on the resulting temperature distributions. This point is discussed further in the section covering the results of steady-state calculations.

Finite difference methods have been used to program for digital computation the ordinary differential equations describing the changes in enthalpy and reactant concentrations in the interstitial phase. No iteration is necessary to solve these equations numerically when the incremental axial distances are sufficiently small. The size of a succeeding increment is calculated at each axial position as a function of the rates of change of temperature and fractional ammonia dissociation with axial distance. However, Eq. (16), which must be solved simultaneously with the differential equations, is an implicit integral equation which requires an iterative procedure for solution. Hand calculations have indicated that convergence to a solution for $c_p(x)$ is difficult to achieve unless the initial estimate of the concentration distribution is fairly accurate. Methods have been developed for generating this estimate and an iterative procedure has been devised which effects

rapid convergence over a fairly wide range of conditions. This procedure is presently used as a subroutine in the main program representing the steady-state model. Sample calculations have been made to illustrate the use of this subroutine for computing concentration profiles in the porous catalyst particles, the associated temperature profiles, and the concentration gradients at the particle surface. The subroutine was run for values of the reaction rate constant, $k_0$, between 2.0 $sec^{-1}$ and 2.0 x $10^6$ $sec^{-1}$, and for $\beta = 0.171$, $\gamma = 18.3$, $D_p = 0.84 \times 10^{-5}$ $ft^2/sec$, $a = 0.005$ ft, $n = 1$, and $(c_p)_s = 0.357$ $lb/ft^3$. The values chosen for $\beta$, $\gamma$, $D_p$, $n$, and $(c_p)_s$ are reasonable for the decomposition of hydrazine within Shell 405 catalyst pellets. The resulting concentration profiles are plotted for each reaction rate constant in Fig. 1 and the associated temperature profiles are shown in Fig. 2. The concentration gradient at the particle surface is plotted as a function of the reaction rate constant in Fig. 3.

Transient Model

In developing the transient model, the temperature and the concentrations of reactants in the interstitial phase are assumed to vary only with time and axial distance along the bed. In this system film coefficients are used to describe heat and mass transfer between the interstitial phase and the outside surface of the catalyst pellets. The reactant concentrations, $c_p$, and the temperature, $T_p$, are taken as uniform within the interior of the porous particles. Heat and mass diffusion within the particles are taken into account by considering their effects on the utilization factor previously defined. That is, reaction rates computed on the basis of uniform $c_p$ and $T_p$ are multiplied by a utilization factor determined by analogy with the steady-state system. In addition, it is assumed that liquid velocities are sufficiently low relative to other rate processes so that, for all practical purposes, steady-state in the liquid and liquid-vapor regions is achieved as soon as the liquid reaches a given axial location in the reactor.

The transient model is concerned then with the vapor region only, where velocities are about three orders of magnitude greater than in the liquid region. Here it is reasonable to assume that gas velocities are so great that the time lag from the entrance to the vapor region to any position z for the fluid is negligible compared with other transient effects. With this assumption the rates of change of temperature and reactant weight fractions with axial distance in the interstitial phase are given by

$$\frac{\partial T_i}{\partial z} = -\frac{H^{N_2H_4}}{G\bar{C}_F} r_{hom}\delta - \frac{A_p}{G\bar{C}_F}\left[h_c\,(T_i - T_p)\right] - \frac{F}{G\bar{C}_F}\left[(h_i^v - h_F) + \bar{C}_F\,(T_i - T_{vap})\right] \qquad (19)$$

$$\frac{\partial w_i^{N_2H_4}}{\partial z} = \frac{1}{G}\left\{F - r_{hom}\,\delta - A_p k_c^{N_2H_4}\left(c_i^{N_2H_4} - c_p^{N_2H_4}\right)\right\} - \frac{Fc_i^{N_2H_4}}{G\rho_i} \qquad (20)$$

$$\frac{\partial w_i^{NH_3}}{\partial z} = \frac{1}{G}\left\{r_{hom}\,\delta\,\frac{M^{NH_3}}{M^{N_2H_4}} - A_p k_c^{NH_3}\left(c_i^{NH_3} - c_p^{NH_3}\right)\right\} - \frac{Fc_i^{NH_3}}{G\rho_i} \qquad (21)$$

$$\frac{\partial w_i^{N_2}}{\partial z} = \frac{1}{G}\left\{r_{hom}\,\delta\,\frac{M^{N_2}}{2M^{N_2H_4}} - A_p\,k_c^{N_2}\left(c_i^{N_2} - c_p^{N_2}\right)\right\} - \frac{Fc_i^{N_2}}{G\rho_i} \qquad (22)$$

$$\frac{\partial w_i^{H_2}}{\partial z} = \frac{1}{G}\left\{r_{hom}\,\delta\,\frac{M^{H_2}}{2M^{N_2H_4}} - A_p\,k_c^{H_2}\left(c_i^{H_2} - c_p^{H_2}\right)\right\} - \frac{Fc_i^{H_2}}{G\rho_i} \qquad (23)$$

The changes in reactant concentrations with axial distance can then be obtained using the above equations together with Eqs. (9), (10), and (11). At a given axial location the rates of change of temperature and reactant concentrations in the catalyst particles with time are given by

$$\frac{dT_p}{dt} = -\frac{1}{\rho_s C_s}\left[\left(H\,r_{het}\right)^{N_2H_4} + \left(H r_{het}\right)^{NH_3}\right] + \frac{3h_c}{a\rho_s C_s}\left(T_i - T_p\right) \qquad (24)$$

$$\frac{dc_p^{N_2H_4}}{dt} = -\frac{1}{\alpha_p}\,r_{het}^{N_2H_4} + \frac{3k_c^{N_2H_4}}{\alpha_p\,a}\left(c_i^{N_2H_4} - c_p^{N_2H_4}\right) \qquad (25)$$

$$\frac{dc_p^{NH_3}}{dt} = \frac{1}{\alpha_p}\left[r_{het}^{N_2H_4}\,\frac{M^{NH_3}}{M^{N_2H_4}} - r_{het}^{NH_3}\right] + \frac{3k_c^{NH_3}}{\alpha_p\,a}\left(c_i^{NH_3} - c_p^{NH_3}\right) \qquad (26)$$

$$\frac{dc_p^{N_2}}{dt} = \frac{1}{\alpha_p}\left[r_{het}^{N_2H_4}\frac{M^{N_2}}{2M^{N_2H_4}} + r_{het}^{NH_3}\frac{M^{N_2}}{2M^{NH_3}}\right] + \frac{3k_c^{N_2}}{\alpha_p a}\left(c_i^{N_2} - c_p^{N_2}\right) \quad (27)$$

$$\frac{dc_p^{H_2}}{dt} = \frac{1}{\alpha_p}\left[r_{het}^{N_2H_4}\frac{M^{H_2}}{2M^{N_2H_4}} + r_{het}^{NH_3}\frac{3M^{H_2}}{2M^{NH_3}}\right] + \frac{3k_c^{H_2}}{\alpha_p a}\left(c_i^{H_2} - c_p^{H_2}\right) \quad (28)$$

Recalling that the reaction of hydrazine on the catalyst surfaces is extremely fast, so that the reaction rate is controlled by the rate of transport of hydrazine to the catalyst surfaces, Eq. (25) can be used to define $r_{het}^{N_2H_4}$ by noting that $(dc_p^{N_2H_4}/dt)$ and $c_p^{N_2H_4}$ are both approximately equal to zero. The reaction rate is then given by

$$r_{het}^{N_2H_4} = \frac{3k_c^{N_2H_4}}{a} c_i^{N_2H_4} \quad (29)$$

The reaction rate of ammonia on the catalyst surfaces, $r_{het}^{NH_3}$, is computed by multiplying the rate of reaction calculated on the basis of uniform $T_p$ and $c_p$ (see section on Kinetics Information) by the utilization factor determined by analogy with the steady-state system. Under steady-state conditions the concentration gradient at the particle surface, $(dc_p^{NH_3}/dx)$, can be used to define this utilization factor which, for a first-order reaction, is given by:

$$\text{UTILIZATION FACTOR} = \left[3D_p(dc_p/dx)\right]_s \Big/ \left[a\,(c_p)_s k_0\right] \quad (30)$$

The subroutine in the steady-state program which calculates concentration distributions in the catalyst particles has been used to compute the concentration gradients at the particle surface and the corresponding utilization factors for about 100 different reactor operating conditions covering the range of interest of each of the parameters in this program. The results were then plotted as a function of a dimensionless diffusion parameter, $a\sqrt{k_0/D_p}$. All of the results fell within the limits of the dashed curves shown in Fig. 4. The mean results are shown as the solid curve in this figure.

Heat and mass transfer coefficients may be estimated from (Ref. 13).

$$k_c^J = \left(\frac{0.61\,G}{\rho_i}\right)\left(\frac{\mu}{\rho_i D_i^J}\right)^{-0.667}\left(\frac{G}{A_p\mu}\right)^{-0.41} \quad (31)$$

$$h_c = 0.74 \left(\frac{G}{A_p \mu}\right)^{-0.41} (\bar{C}_F G) \tag{32}$$

The ordinary and partial differential equations representing the transient model of the reactor system (Eqs. (19) through (28) along with Eqs. (9) through (11)) must be solved simultaneously for the variation of temperatures and reactant concentrations with time and axial position in the reactor. The partial differential equations representing the changes in interstitial temperature and reactant concentrations with axial distance can be treated as ordinary differential equations by integrating them at constant time. This can be accomplished, together with integrating Eqs. (24) through (28) at fixed axial positions, by establishing a network containing fixed time intervals and intervals of axial distance. Ideally, then, reactant concentrations and temperatures in the interstitial phase and in the catalyst particles could be computed at the $i$th time and $j$th axial distance from

$$(T_p)_{i,j} = (T_p)_{i-1,j} + \left(\frac{dT_p}{dt}\right)_{i-1,j} \Delta t \tag{33}$$

$$(c_p)^J_{i,j} = (c_p)^J_{i-1,j} + \left(\frac{dc_p}{dt}\right)^J_{i-1,j} \Delta t \tag{34}$$

$$(T_i)_{i,j} = (T_i)_{i,j-1} + \left(\frac{\partial T_i}{\partial z}\right)_{i,j-1} \Delta z \tag{35}$$

$$(c_i)^J_{i,j} = (c_i)^J_{i,j-1} + \left(\frac{\partial c_i}{\partial z}\right)^J_{i,j-1} \Delta z \tag{36}$$

This procedure is satisfactory as long as the intervals of time and axial distance are chosen to be sufficiently small so that the differential quantities are approximately constant in each interval. In the system considered here this restriction leads to rather long computing times. The restriction can be relaxed somewhat by rearranging each of the differential equations in the form

$$\frac{dg}{ds} = \alpha - \beta g \tag{37}$$

where it is required now that only the quantities $\alpha$ and $\beta$ remain approximately constant while integrating the equation from $s_{k-1}$ to $s_k$ (corresponding to $\Delta t$ or $\Delta z$ ). Equation (37) can be integrated to obtain

$$g_k = g_{k-1} e^{-\beta(s_k - s_{k-1})} + \frac{\alpha}{\beta}\left[1 - e^{-\beta(s_k - s_{k-1})}\right] \qquad (38)$$

where $g_k$ is the value of g at $s_k$ , and $g_{k-1}$ is the value of g at $s_{k-1}$ . An alternative form of Eq. (38) is

$$g_k = g_{k-1} + \left(\frac{dg}{ds}\right)_{k-1}\left[\frac{1 - e^{-\beta(s_k - s_{k-1})}}{\beta}\right] \qquad (39)$$

It can be seen that, for small values of ( $s_k - s_{k-1}$ ), Eq. (39) reduces to the same form as Eqs. (33) through (36). It is convenient to use equations of the form of Eq. (38) in place of Eqs. (33) and (34) to compute particle concentrations and temperatures, and to use equations of the form of Eq. (39) in place of Eqs. (35) and (36) to compute interstitial concentrations and temperatures.

Kinetics Information

A number of investigators have reported rate expressions for the kinetics of homogeneous, vapor-phase decomposition of hydrazine (Refs. 7, 8, and 9). The rate expression chosen for use in this work was that of Ref. 7 because the results were obtained in a reactor similar to the one of interest here. The equation reported is

$$r_{hom}^{N_2H_4} = 2.14 \times 10^{10}\ c_i^{N_2H_4}\ e^{-33{,}000/T_i} \qquad lb/ft^3 - sec \qquad (40)$$

where $c_i^{N_2H_4}$ is in $lb/ft^3$ and $T_i$ is in deg R.

Similar information relating to the catalytic decomposition of hydrazine is not available. However, a rate expression of the same form as Eq. (40) was chosen to represent the kinetics of the heterogeneous decomposition. Qualitative evidence from rocket firings at various feed temperatures indicates a rather low activation energy. A value of 2500 x 1.986 Btu/lb-mole was chosen rather arbitrarily and a preexponential factor in the Arrhenius type rate law was computed by using engine test data to make a rough estimate of the reaction rate at a typical feed temperature. These approximations have led to the following expression for reaction rate:

$$r_{het}^{N_2H_4} = 10^{10}\ c_p^{N_2H_4}\ e^{-2500/T_p} \qquad lb/ft^3 - sec \qquad (41)$$

where $c_p^{N_2H_4}$ is in $lb/ft^3$ and $T_p$ is in deg R.

Melton (Ref. 14) and Logan and Kemball (Ref. 15) have conducted low pressure experiments on the vapor-phase decomposition of ammonia on platinum catalysts. They report a reaction rate expression of the form:

$$r_{het}^{NH_3} = \alpha^{NH_3} \frac{C_P^{NH_3}}{\left[C_P^{H_2}\right]^n} e^{-50{,}000/T_p} \qquad lb/ft^3 - sec \qquad (42)$$

where $T_p$ is in deg R and n is given as 1.0 by Melton and as 1.6 by Logan and Kemball. Melton states however that ". . . The inhibitive action of hydrogen is probably a little greater than that corresponding to the first power of hydrogen. . ." A similar dependence on hydrogen concentration at higher pressures was questionable particularly if hydrogen was physically absorbed on the platinum catalyst. Increasing hydrogen concentration beyond a certain point might then have much less effect on impeding the reaction rate. It was decided to perform computer calculations using the steady-state program assuming no dependence of $r_{het}^{NH_3}$ on the hydrogen concentration and then see if the value of $\alpha^{NH_3}$ which best fits the experimental data decreases as the pressure increases. Such behavior would indicate that the dependence on hydrogen concentration did indeed extend to higher pressures. This behavior was in fact observed. It was assumed, of course, that the order of the ammonia dissociation reaction with respect to hydrogen is the same on the Shell 405 catalyst as it is on platinum. Although this assumption remains untested, subsequent calculations using the steady-state program exhibit the best agreement with the available engine test data when $r_{het}^{NH_3}$ is given by

$$r_{het} = 0.3 \times 10^{11} \frac{C_p^{NH_3}}{(C_p^{H_2})^{1.6}} e^{-50{,}000/T_p} \qquad lb/ft^3 - sec \qquad (43)$$

where the concentrations are expressed in $lb/ft^3$.

It should be emphasized that the value of $0.3 \times 10^{11}$ for $\alpha^{NH_3}$ was obtained by comparing experimental data with temperature and concentration profiles calculated using the computer program representing the steady-state model. It is implicitly assumed in this model that the temperature at the surface of the catalyst pellet is equal to the interstitial temperature at the same axial location. This assumption has little effect on the calculated rates of the diffusion-controlled hydrazine decomposition and it is fairly accurate in regions of the reaction chamber where only ammonia is decomposing. However, in regions where both hydrazine and ammonia are decomposing simultaneously, the catalyst particle surface temperatures may be higher than the interstitial temperatures. If this were taken into account in the steady-state model the value of $\alpha^{NH_3}$ exhibiting closest agreement with experimental data would be somewhat larger than $0.3 \times 10^{11}$. In the simpler format of the transient model the differences between the particle surface and interstitial temperatures are accounted for and the rates of heat transfer corresponding to these temperature differences are expressed in terms of a heat transfer coefficient.

Theoretically then, the transient model, when extrapolated to steady-state and compared with experimental data, should yield a more accurate value of $\alpha^{NH_3}$. The value obtained using the extrapolated transient results is $1 \times 10^{11}$. However, heat transfer coefficients used in this model are estimated using correlations developed for nonreacting systems. In regions where significant chemical reaction takes place, such as the zones in which both hydrazine and ammonia are decomposing, actual heat transfer rates are considerably higher than those calculated in the transient model. This would lead to a value of $\alpha^{NH_3}$ less than $1 \times 10^{11}$. The actual value of $\alpha^{NH_3}$ is probably between the one estimated using the steady-state program ($0.3 \times 10^{11}$) and the one estimated using the transient program ($1 \times 10^{11}$).

## Results of Calculations

Calculations were made with the computer programs representing the steady-state and transient models of the reactor system to evaluate the effects on temperature and reactant concentration distributions of such parameters as initial bed temperature, feed temperature, chamber pressure, mass flow rate, axial injection profile, and catalyst size distribution. The sensitivity of the calculated results to variations in the rate constants associated with the catalytic decomposition reactions was also determined. The results have been compared, wherever possible, to available measurements of temperature profiles and exhaust concentrations made during firings of typical hydrazine engine reaction chambers.

### Steady-State Results

A series of calculations was made using the steady-state program for cases for which experimental information is available (Ref. 2) in order to examine the effectiveness of the steady-state model and to establish the rate constants associated with the catalytic decomposition reactions. The results of these calculations are shown in Figs. 5 through 19. For each of these cases the catalyst bed packing was taken to consist of 25-30 mesh catalyst particles for the first 0.2 in. and 1/8 in. x 1/8 in. cylindrical pellets for the remainder of the bed. This configuration is referred to in the figures as the "standard bed configuration." Except where the rate constants $\alpha^{NH_3}$ and $\alpha^{N_2H_4}$ were varied, the rates of the catalytic decomposition reactors were given by Eqs. (41) and (43). Equation (43) was simplified somewhat by noting that, since hydrogen diffusion is quite rapid, $c_p^{H_2} \sim c_i^{H_2}$. The reactor operating conditions of chamber pressure and mass flow rate were varied, as was the feed temperature, $T_F$; although, in each of these cases, $T_F$ was between approximately 510 and 530 deg R.

Temperature distributions are plotted in Fig. 5 for a case in which the reactor operating conditions were taken as $G = 3.12$ lb/ft$^2$- sec and $P = 479.5$ psia. Temperature profiles are plotted for $\alpha^{NH_3} = 0.3 \times 10^{10}$, $0.3 \times 10^{11}$, and $0.3 \times 10^{12}$ (lb/ft$^3$)$^{0.6}$ (sec)$^{-1}$. An enlarged view of the temperature profile in the liquid-vapor region is shown to illustrate the effects of pure liquid as opposed to a combination of liquid and vapor wetting the outside surface of the catalyst particles. Also shown in Fig. 5 are temperature measurements obtained by Rocket Research

Corporation (Ref. 2) during the course of engine firings under the same operating conditions. Comparison of the calculated results with the experimental data indicates better correspondence of results with $\alpha^{NH_3} = 0.3 \times 10^{11}$ $(lb/ft^3)^{0.6}$ $(sec)^{-1}$. The calculated mole-fraction profiles associated with this rate constant, together with experimental values of mole-fractions, are shown in Fig. 6. The corresponding fractional ammonia dissociation profile is shown in Fig. 7. The results of similar calculations made for G = 2.43 $lb/ft^2$-sec and P = 1042 psia are shown in Figs. 8 through 10; those made for G = 6.29 $lb/ft^2$-sec and P = 974 psia are shown in Figs. 11 through 13; those made for G = 1.52 $lb/ft^2$-sec and P = 217.9 psia are shown in Figs. 14 through 16, and those made for G = 1.51 $lb/ft^2$-sec and P = 111.4 psia are shown in Figs. 17 through 19.

In each of the above calculations the rate constant for the catalytic decomposition of hydrazine was taken as $10^{10}$ $sec^{-1}$. Since catalytic decomposition of hydrazine is diffusion-controlled in the vapor region, varying this rate constant would only affect the width of the liquid and liquid-vapor regions. These regions are extremely narrow for the cases examined above because of the large surface area of the small catalyst particles in these regions and because of the high vapor pressure of liquid hydrazine at the feed temperatures considered.

The effects of various reactor operating conditions on steady-state temperature profiles are illustrated more specifically in Figs. 20 through 23. A reference case was chosen in which the operating conditions were G = 3.0 $lb/ft^2$-sec, P = 100 psia, and $T_F$ = 530 R, and the catalyst bed configuration was the "standard bed configuration." Chamber pressure, mass flow rate, catalyst bed configuration and axial injection profile were then varied in turn and the resulting temperature distributions were plotted. In Fig. 20 temperature distributions are plotted for chamber pressures of 100, 500, and 1000 psia with all other conditions taken as those of the reference case. Increasing pressure causes the peak temperature to rise and shift slightly toward the inlet of the reactor. This is due to the inhibiting action of hydrazine on the rate of ammonia dissociation. Temperature profiles are shown in Fig. 21 for mass flow rates of 1.5, 3.0, and 6.0 $lb/ft^2$-sec. Increasing flow rate causes the peak temperature to rise and shift slightly away from the reactor inlet. The effect of changing the catalyst bed configuration on the temperature distribution is shown in Fig. 22. Temperature profiles are plotted for beds packed with all 25-30 mesh particles, all 1/8 in. x 1/8 in. cylindrical pellets, and the standard non-uniform particle size distribution. It is apparent that the larger particles slow down the rates of the catalytic decomposition reactions. The effects of distributed injectors on temperature profiles are illustrated in Fig. 23. Temperature distributions are plotted for the reference case, for the case in which 2/3 of the hydrazine is injected at the inlet and the remaining 1/3 is injected uniformly over the first 1/2 in. of the reactor, and for the case in which 1/3 of the hydrazine is injected at the inlet and the remaining 2/3 is injected uniformly over the first 1/2 in. of the reactor. In these cases the bed was taken to be packed with all 25-30 mesh particles.

Steady-state temperature and reactant concentration distributions can also be obtained using the transient program by running the calculations out to long (essentially infinite) times. A comparison of the steady-state axial temperature profiles computed from the steady-state and transient programs is shown in Fig. 24 for reactor operating conditions of $G = 3.12$ lb/ft$^2$- sec, P(feed pressure) = 479.5 psia, and $T_F = 509$ deg R and for the standard bed configuration. The rate constant for the catalytic decomposition of ammonia, $\alpha^{NH_3}$, was taken as $10^{11}$ (lb/ft$^3$)$^{0.6}$ (sec)$^{-1}$ (see section on Kinetics Information). A comparison of the associated axial profiles of mole-fraction of ammonia computed from the steady-state and transient models is shown in Fig. 25.

Transient Results

A series of calculations was made using the transient program for the same cases cited previously in order to examine the effectiveness of the transient model of the reactor system. The results of these calculations are shown in Figs. 26 through 44. In each of these cases the catalyst bed configuration was taken as the standard configuration, the feed temperature varied slightly between approximately 510 and 530 deg R, and the initial bed temperature was taken equal to the feed temperature. Temperature distributions are plotted for various times up to steady-state in Fig. 26 for a case in which the reactor operating conditions were taken as $G = 3.12$ lb/ft$^2$- sec and P(feed pressure) = 479.5 psia. The vaporization time noted in Fig. 26 is defined as the time required, under steady-state conditions, for liquid hydrazine to flow from the reactor inlet to the interface between the liquid-vapor and vapor regions. Temperatures are cross-plotted as a function of time at three axial positions in Fig. 27. Also shown in Fig. 27 are temperature measurements obtained by Rocket Research Corporation (Ref. 16) during the course of engine firings under the same operating conditions. The calculated mole-fraction profiles corresponding to the temperature profiles shown in Fig. 26 are illustrated in Figs. 28, 29, 30 and 31 for hydrazine, ammonia, nitrogen, and hydrogen, respectively. The chamber exit pressure for this case was calculated as a function of time and plotted in Fig. 32. It is interesting to note that the chamber exit pressure rises sharply, initially, as hydrazine and the products of hydrazine decomposition in the upstream end of the reaction chamber flow through the reactor, drops as hydrazine decomposes on the catalyst surfaces further downstream in the reactor and as the products of hydrazine decomposition diffuse into the catalyst particles, and rises again once the concentrations of the products of hydrazine decomposition equilibrate between the vapor phase and the catalyst particles in the downstream end of the reactor. The results of similar calculations made for $G = 2.43$ lb/ft$^2$-sec and P (feed pressure) = 1040 psia are shown, except for the concentration profiles, in Figs. 33 through 35; those made for $G = 6.29$ lb/ft$^2$-sec and $P = 974$ psia are shown in Figs. 36 through 38; those made for $G = 1.52$ lb/ft$^2$-sec and $P = 217.9$ psia are shown in Figs. 39 through 41, and those made for $G = 1.51$ lb/ft$^2$-sec and $P = 111.4$ psia are shown in Figs. 42 through 44.

The effects of various reactor operating conditions on the variation of gas temperature at the chamber exit with time can be better illustrated by comparing, with a reference case, the results of calculations made in turn for various chamber pressures, mass flow rates, catalyst bed configurations, axial injection profiles, and initial bed temperatures. A reference case was chosen in which the operating conditions were taken as G = 3.0 $lb/ft^2$- sec, P (feed pressure) = 100 psia, $T_F$ = 530 deg R and $T_p$ (initial) = 530 deg R, and the catalyst bed configuration was taken as the "standard bed configuration." The bed length was taken as 0.25 ft. In Fig. 45 exit gas temperatures are plotted as a function of time for feed pressures of 100 and 1000 psia with all other conditions taken as those of the reference case. Very little effect of pressure on transient response is noted. The marked effect of mass flow rate on transient response time is shown in Fig. 46 where exit gas temperatures are plotted versus time for mass flow rates of 1.5, 3.0, and 6.0 $lb/ft^2$- sec. The effect of changing the catalyst bed configuration on the temperature variation with time is illustrated in Fig. 47. Exit gas temperatures are plotted as a function of time for beds packed with all 25-30 mesh particles, all 1/8 in. x 1/8 in. cylindrical pellets, and the standard non-uniform particle size distribution. The effects of distributed injectors on temperature profiles is shown in Fig. 48. Exit gas temperatures are plotted as functions of time for a constant mass flow rate and for the case in which 1/3 of the hydrazine is injected at the inlet and the remaining 2/3 is injected uniformly over the first 1/2 in. of the reactor. In these cases the bed was taken to be packed with all 25-30 mesh particles and the initial bed temperature was taken as 1200 deg R in order to insure vaporization of the cold hydrazine injected downstream of the inlet. Finally, the effects of varying the initial bed temperatures are illustrated in Fig. 49. Exit gas temperatures are plotted versus time in Fig. 49 for initial bed temperatures of 530, 1200, and 1500 deg R.

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## LIST OF SYMBOLS

| | |
|---|---|
| $a$ | Radius of spherical particle, ft |
| $A_p$ | Total external surface of catalyst particle per unit volume of bed, $ft^{-1}$ |
| $c_i$ | Reactant concentration in interstitial fluid, $lb/ft^3$ |
| $c_p$ | Reactant concentration in gas phase within the porous particle, $lb/ft^3$ |
| $c_p^*$ | Equals $c_p - (c_p)_s$ , $lb/ft^3$ |
| $C_f$ | Specific heat of fluid in the interstitial phase, Btu/lb - deg R |
| $\bar{C}_f$ | Average specific heat of fluid in the interstitial phase, Btu/lb - deg R |
| $C_s$ | Specific heat of catalyst particle, Btu/lb - deg R |
| $D_i$ | Diffusion coefficient of reactant gas in the interstitial fluid, $ft^2/sec$ |
| $D_p$ | Diffusion coefficient of reactant gas in the porous particle, $ft^2/sec$ |
| $F$ | Rate of feed of hydrazine from distributed injectors into the system, lb/ft3- sec |
| $g$ | Dependent variable in Eq. (37) |
| $G$ | Mass flow rate, $lb/ft^2$- sec |
| $h$ | Enthalpy, Btu/lb |
| $h_c$ | Heat transfer coefficient, $Btu/ft^2$- sec - deg R |
| $H$ | Heat of reaction (negative for exothermic reaction), Btu/lb |
| $k_c$ | Mass transfer coefficient, ft/sec |
| $k_o$ | Reaction rate constant, equals $\alpha e^{-\gamma}$ , $sec^{-1}$ |
| $K_p$ | Thermal conductivity of the porous catalyst particle, Btu/ft-sec-deg R |
| $M$ | Molecular weight, lb/lb mole |
| $\bar{M}$ | Average molecular weight, lb/lb mole |
| $n$ | Order of decomposition reaction |
| $P$ | Chamber pressure, psia |

LIST OF SYMBOLS (cont'd)

| | |
|---|---|
| $Q$ | Activation energy, Btu/lb mole |
| $r_{het}$ | Rate of (heterogeneous) chemical reaction on the catalyst surfaces, $lb/ft^3$ - sec |
| $r_{hom}$ | Rate of (homogeneous) chemical reaction in the interstitial phase, $lb/ft^3$ - sec |
| $s$ | Independent variable in Eq. (37) |
| $R$ | Gas constant, equals 10.73 psia - $ft^3$/lb mole - deg R |
| $t$ | Time, sec |
| $t^*$ | Actual time minus time required, under steady-state conditions, for liquid hydrazine to flow from the reactor inlet to the interface between the liquid-vapor and vapor regions, sec |
| $T$ | Temperature, deg R |
| $u$ | Mathematical function (defined in Appendix I) |
| $v$ | Mathematical function (defined in Appendix I) |
| $w_i$ | Weight fraction of reactant in interstitial phase |
| $x$ | Radial distance from the center of the spherical catalyst particle, ft |
| $z$ | Axial distance, ft |
| $\alpha$ | Preexponential factor in rate equation, or<br>Constant in Eq. (37) |
| $\alpha_p$ | Intraparticle void fraction |
| $\beta$ | Equals $\frac{-(c_p)_s \, H \, D_p}{K_p \, (T_p)_s}$ , or<br>Constant in Eq. (37) |
| $\gamma$ | Equals $Q / R(T_p)_s$ |
| $\delta$ | Interparticle void fraction |
| $\mu$ | Viscosity of interstitial fluid, lb/ft - sec |
| $\rho_i$ | Density of interstitial fluid, $lb/ft^3$ |
| $\rho_s$ | Bulk density of catalyst particle, $lb/ft^3$ |

LIST OF SYMBOLS (cont'd)

| | |
|---|---|
| $\phi$ | Mathematical function (defined in Appendix I) |

Subscripts

| | |
|---|---|
| F | Refers to feed |
| i | Refers to interstitial phase, or |
| | Time index in transient model |
| j | Axial distance index in transient model |
| k | Index in Eqs. (38) and (39) |
| P | Refers to gas within the porous catalyst particle |
| S | Refers to surface of catalyst particle |

Superscripts

| | |
|---|---|
| J | Refers to chemical species |
| L | Refers to liquid at vaporization temperature |
| V | Refers to vapor at vaporization temperature |

# APPENDIX I

## DERIVATION OF INTEGRAL EQUATIONS REPRESENTING THE CONCENTRATION PROFILES OF REACTANTS WITHIN THE CATALYST PARTICLES

In this section equations are developed to describe the steady-state concentration profiles of hydrazine vapor and of ammonia within the catalyst particles. The reactant concentration profiles in the porous particles at any axial location can be found as solutions to:

$$D_p \nabla^2 c_p - r_{het}(c_p) = 0 \qquad \text{(I-1)}$$

If the catalyst particles are taken to be "equivalent" spheres of radii $a$, and if concentration $c_p^*$ is defined such that $c_p^* = c_p - (c_p)_s$, Eq. (I-1) can be written as

$$D_p \left[ \frac{1}{x^2} \frac{d}{dx} \left( x^2 \frac{dc_p^*}{dx} \right) \right] - r_{het} = 0 \qquad \text{(I-2)}$$

where $x$ is the radial distance from the center of a sphere. The boundary conditions associated with Eq. (I-2) are

$$c_p^* = 0 \quad \text{AT} \quad x = a, \qquad \frac{dc_p^*}{dx} = 0 \quad \text{AT} \quad x = 0 \qquad \text{(I-3)}$$

Equation (I-2) can be rearranged to get

$$\frac{d}{dx} \left( x^2 \frac{dc_p^*}{dx} \right) = \frac{r_{het} x^2}{D_p} = \phi(x, c_p^*) \qquad \text{(I-4)}$$

The solution to Eq. (I-4) is most easily obtained by converting it into a Fredholm integral equation (see Ref. 17) of the form

$$c_p^*(x) = \frac{1}{x^2 \left[ u(x) v'(x) - u'(x) v(x) \right]} \int_0^a G(x, \xi) \, \phi(\xi, c_p^*) \, d\xi \qquad \text{(I-5)}$$

where $u(x)$ is a solution of

$$\frac{d}{dx}\left(x^2 \frac{du}{dx}\right) = 0 \tag{I-6}$$

subject to the condition that

$$\left[u \frac{dc_p^*}{dx} - \frac{du}{dx} c_p^*\right]_{x=0} = 0 \tag{I-7}$$

and $v(x)$ is a solution of

$$\frac{d}{dx}\left(x^2 \frac{dv}{dx}\right) = 0 \tag{I-8}$$

subject to the condition that

$$\left[v \frac{dc_p^*}{dx} - \frac{dv}{dx} c_p^*\right]_{x=a} = 0 \tag{I-9}$$

The Green's function, $G(x,\xi)$ is given by

$$G(x,\xi) = \begin{cases} u(\xi)\, v(x) & \text{FOR} \quad 0 \le \xi \le x \\ u(x)\, v(\xi) & \text{FOR} \quad x \le \xi \le a \end{cases} \tag{I-10}$$

The function $u(x)$ can be determined by first integrating Eq. (I-6) to get

$$u = -\frac{A_1}{x} + B_1 \tag{I-11}$$

Applying Eq. (I-7) together with the first of boundary conditions (I-3) to Eq. (I-11), it is found that $A = 0$ and

$$u = B_1 \tag{I-12}$$

The function $v(x)$ can be determined in a similar manner by first integrating Eq. (I-8) to get

$$v = -\frac{A_2}{x} + B_2 \qquad \text{(I-13)}$$

and then applying Eq. (I-9) and the second of boundary conditions (I-3) to Eq. (I-13) to get

$$v = A_2\left[\frac{1}{a} - \frac{1}{x}\right] \qquad \text{(I-14)}$$

Equations (I-10), (I-12), and (I-14) can now be combined to get

$$G(x,\xi) = \begin{cases} A_2 B_1\left[\frac{1}{a} - \frac{1}{x}\right] & \text{FOR} \quad 0 \le \xi \le x \\ A_2 B_1\left[\frac{1}{a} - \frac{1}{\xi}\right] & \text{FOR} \quad x \le \xi \le a \end{cases} \qquad \text{(I-15)}$$

In addition,

$$x^2\left[u(x)\,v'(x) - u'(x)\,v(x)\right] = A_2 B_1 \qquad \text{(I-16)}$$

Equations (I-15) and (I-16) can now be substituted into Eq. (I-5) to get

$$c_p^*(x) = \left[\frac{1}{a} - \frac{1}{x}\right]\int_0^x \phi(\xi, c_p^*)\,d\xi + \int_x^a \left[\frac{1}{a} - \frac{1}{\xi}\right]\phi(\xi, c_p^*)\,d\xi \qquad \text{(I-17)}$$

or

$$c_p(x) = (c_p)_s - \left[\frac{1}{x} - \frac{1}{a}\right]\int_0^x \xi^2 \frac{r_{het}(c_p)}{D_p}\,d\xi - \int_x^a \left[\frac{1}{\xi} - \frac{1}{a}\right]\xi^2 \frac{r_{het}(c_p)}{D_p}\,d\xi \qquad \text{(I-18)}$$

Equation (I-18) is an implicit integral equation which can be solved numerically to determine the concentration at any point in a porous particle in terms of $(c_p)_s$, the concentration at the surface of the particle. The gradients of $c_p$, evaluated at the particle surfaces at given axial locations, can then be obtained from the slopes of curves of $c_p$ versus $x$ at $x = a$.

FIG. 1

EFFECT OF REACTION RATE CONSTANT
ON CONCENTRATION PROFILE WITHIN CATALYST PARTICLE

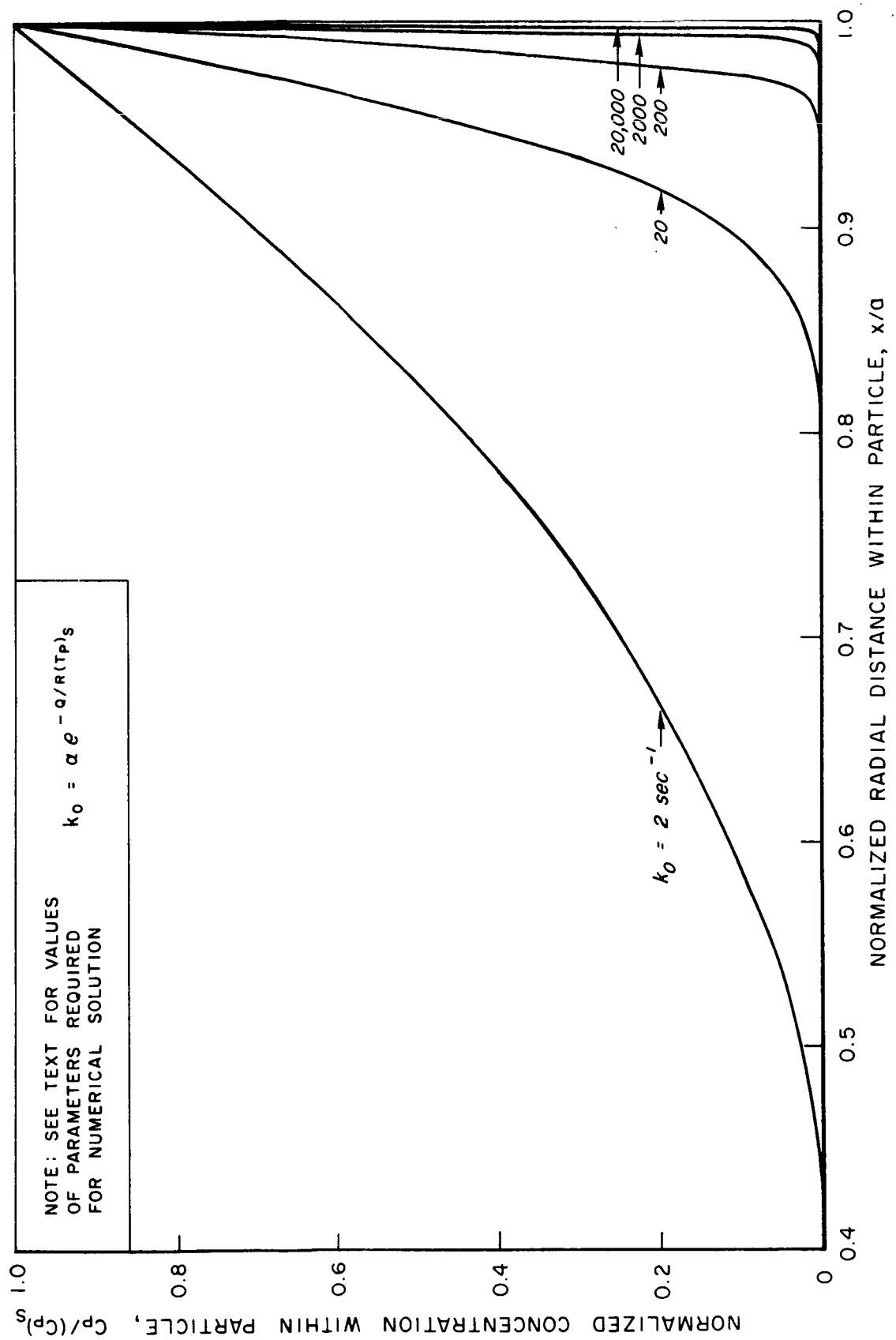


FIG. 2

EFFECT OF REACTION RATE CONSTANT
ON TEMPERATURE PROFILE WITHIN CATALYST PARTICLE

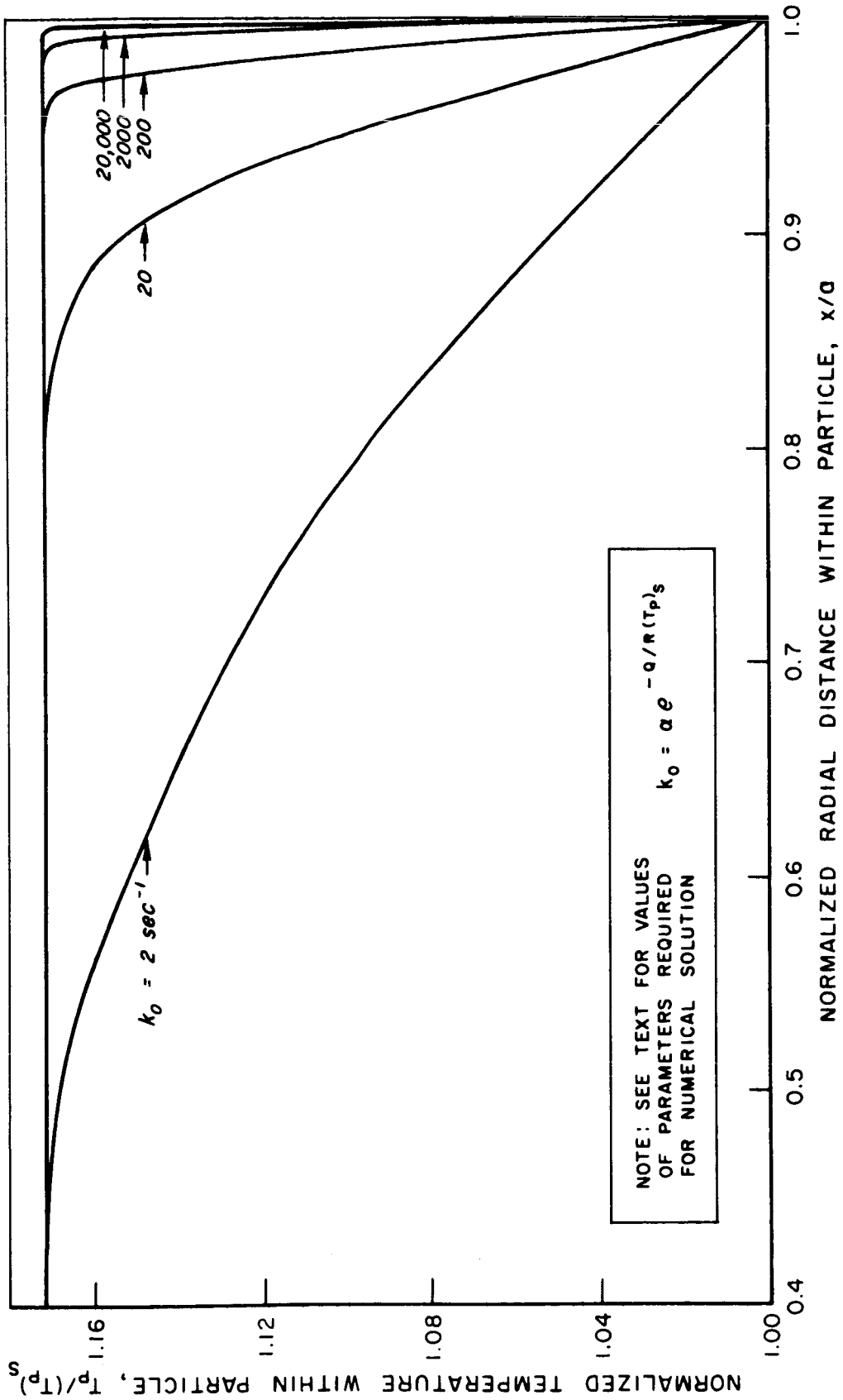


FIG. 3

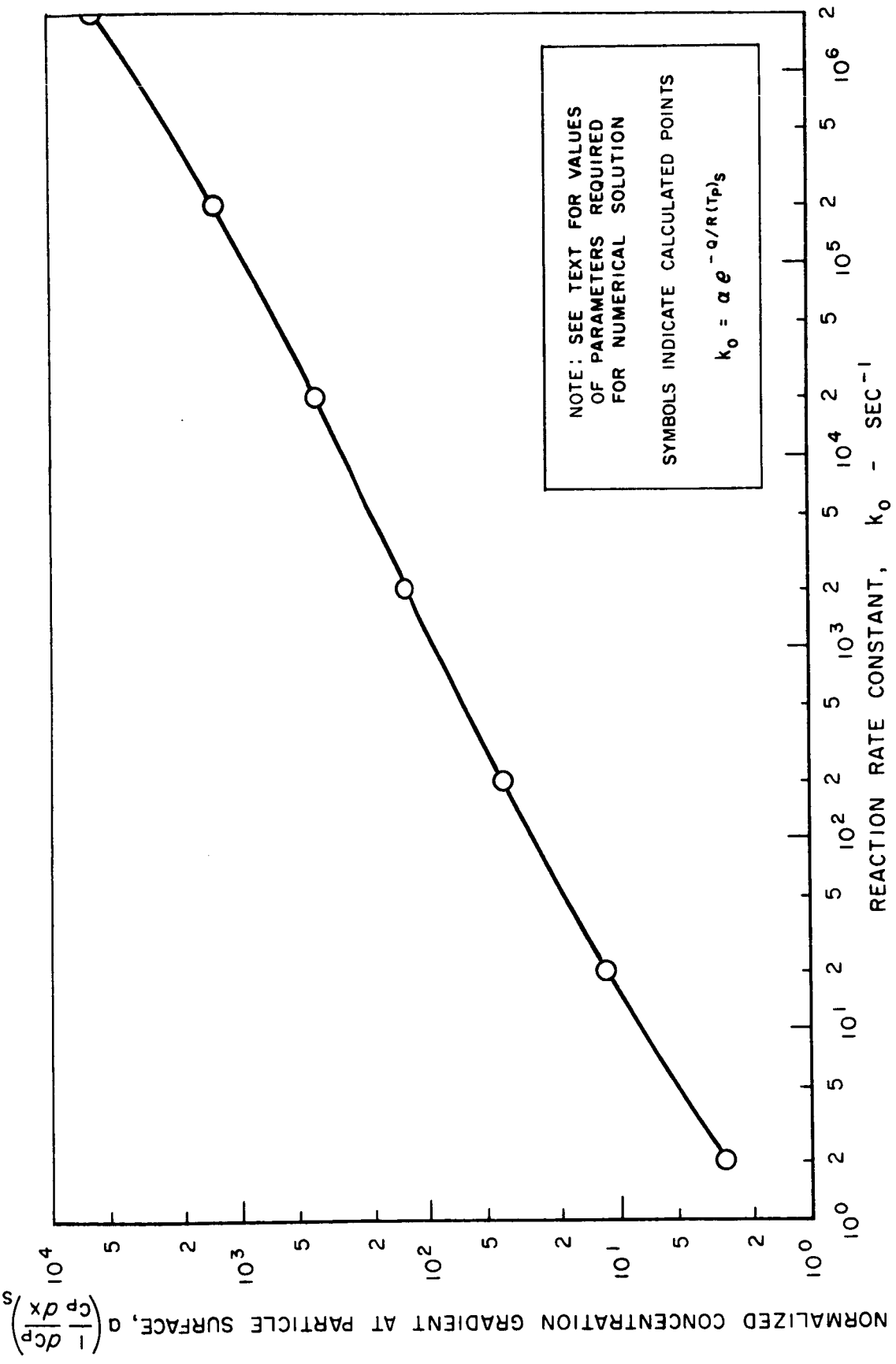

VARIATION OF CONCENTRATION GRADIENT AT PARTICLE SURFACE WITH REACTION RATE CONSTANT
NOTE: SEE TEXT FOR VALUES OF PARAMETERS REQUIRED FOR NUMERICAL SOLUTION
SYMBOLS INDICATE CALCULATED POINTS
$k_o = \alpha e^{-Q/R(T_p)_s}$
NORMALIZED CONCENTRATION GRADIENT AT PARTICLE SURFACE, $\alpha \left(\frac{1}{C_p}\frac{dC_p}{dx}\right)_s$
REACTION RATE CONSTANT, $k_o$ - $SEC^{-1}$
$10^0$ 2 5 $10^1$ 2 5 $10^2$ 2 5 $10^3$ 2 5 $10^4$ 2 5 $10^5$ 2 5 $10^6$ 2
$10^0$ 2 5 $10^1$ 2 5 $10^2$ 2 5 $10^3$ 2 5 $10^4$

FIG. 4

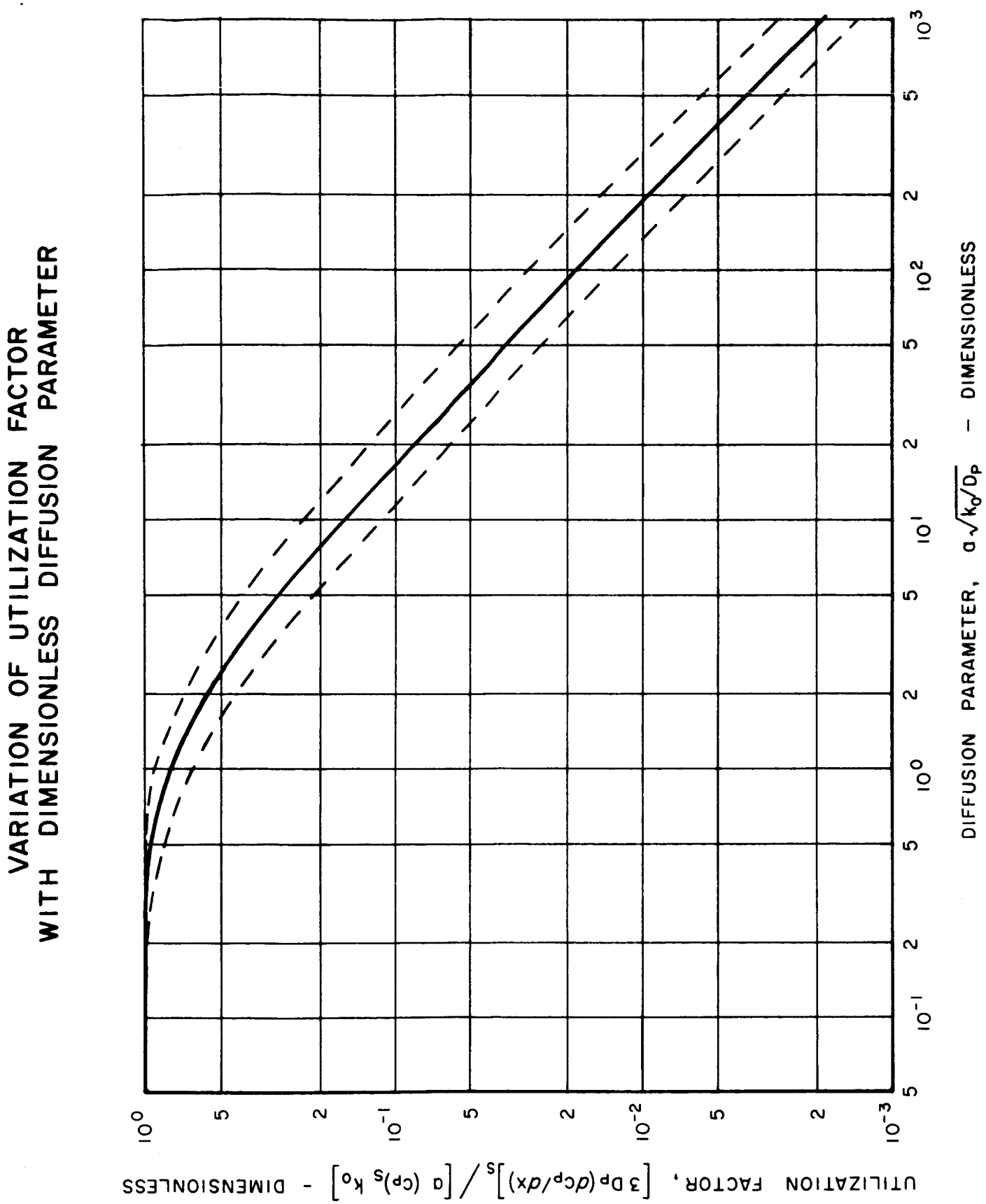

VARIATION OF UTILIZATION FACTOR
WITH DIMENSIONLESS DIFFUSION PARAMETER
UTILIZATION FACTOR, $[3 D_P (dc_P/dx)]_S / [a (cp)_S k_o]$ - DIMENSIONLESS
DIFFUSION PARAMETER, $a\sqrt{k_o/D_P}$ - DIMENSIONLESS
$10^0$
5
2
$10^{-1}$
5
2
$10^{-2}$
5
2
$10^{-3}$
5
$10^{-1}$
2
5
$10^0$
2
5
$10^1$
2
5
$10^2$
2
5
$10^3$

FIG. 5

# STEADY-STATE AXIAL TEMPERATURE PROFILES FOR VARIOUS RATE CONSTANTS FOR THE CATALYTIC DECOMPOSITION OF AMMONIA

P = 479.5 PSIA
G = 3.12 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

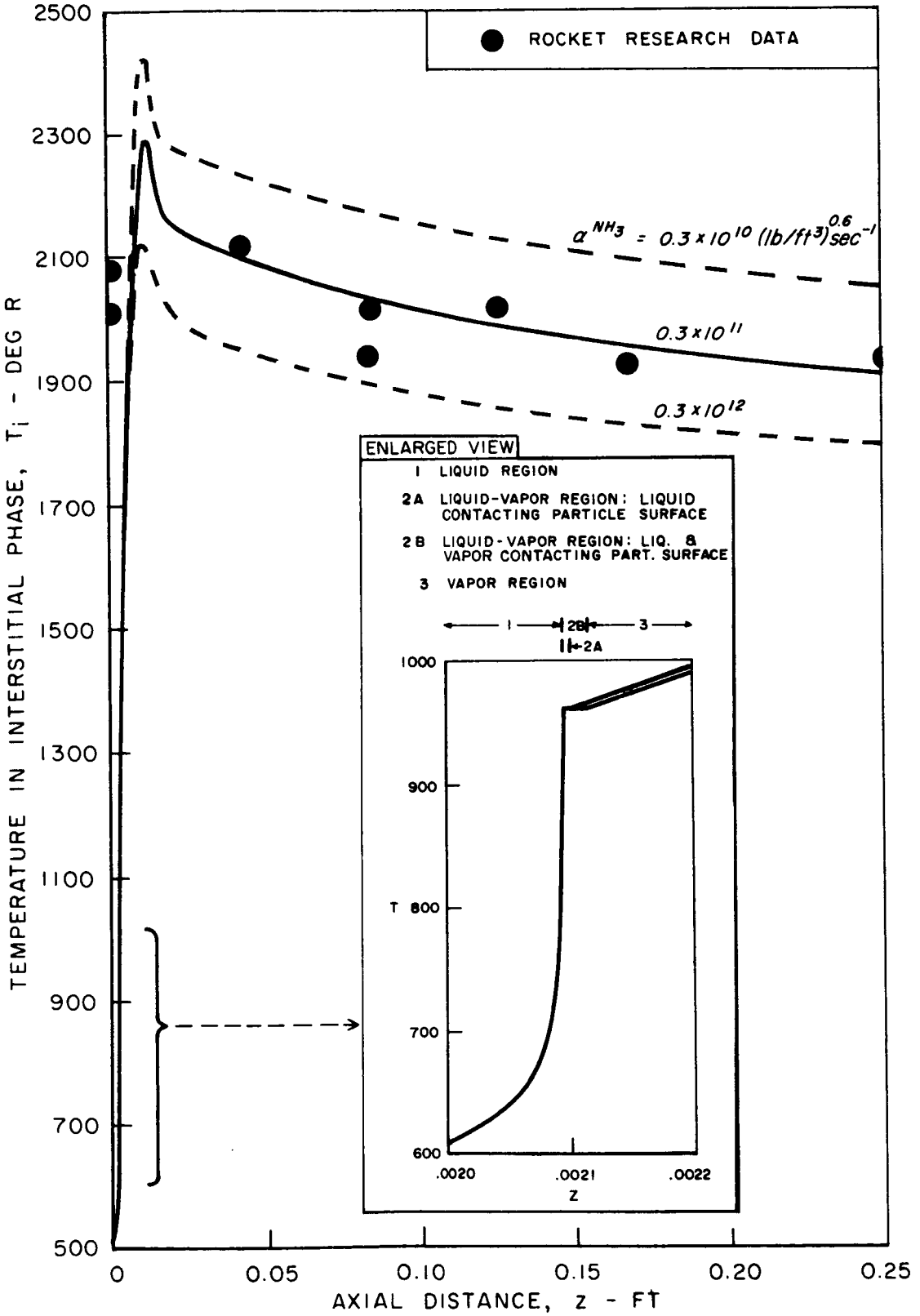


FIG. 6

# STEADY-STATE AXIAL PROFILES OF MOLE-FRACTIONS OF REACTANTS

P = 479.5 PSIA
G = 3.12 LB/FT$^2$-SEC
STANDARD BED CONFIGURATION (SEE TEXT)

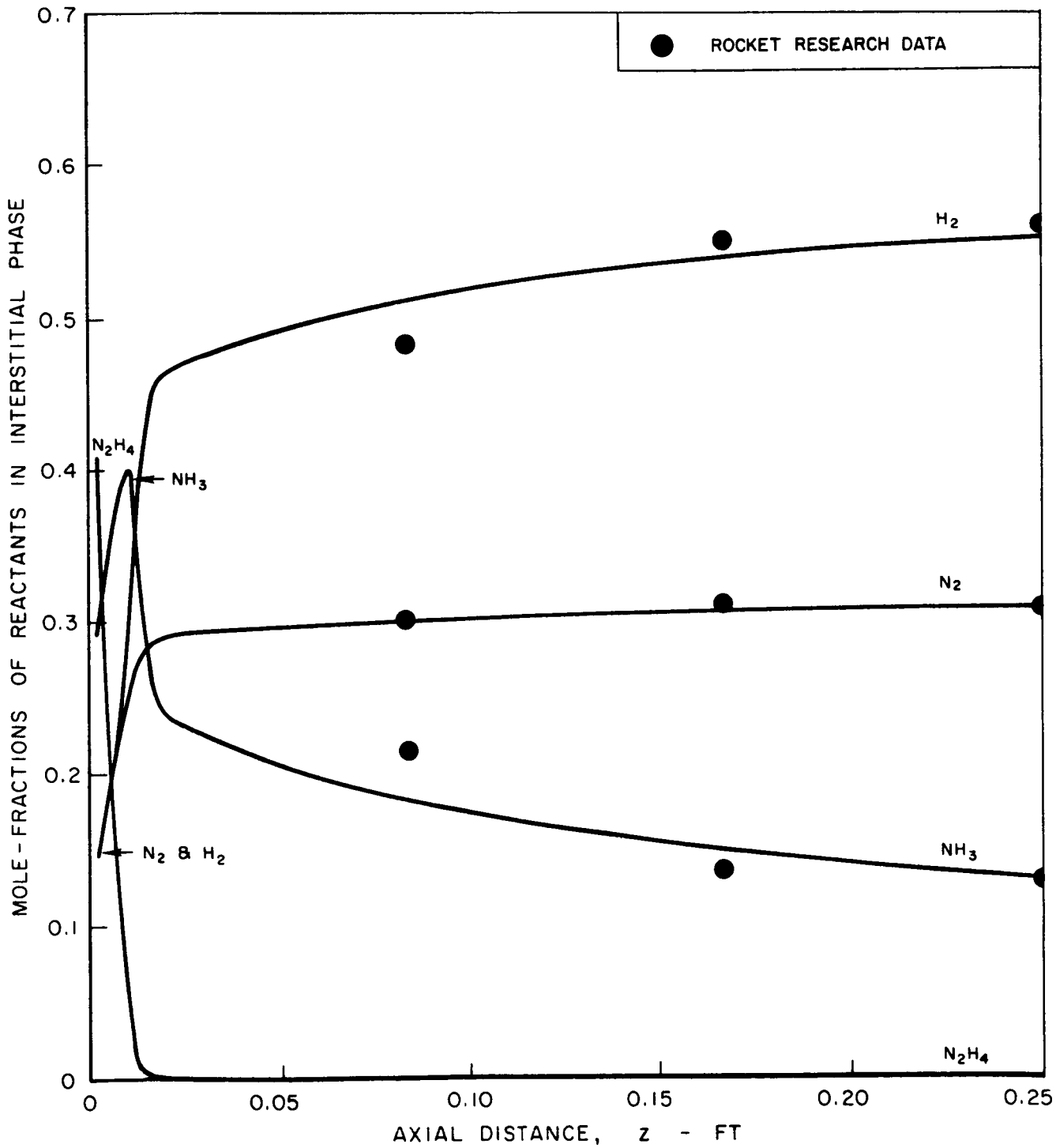


FIG. 7

# STEADY-STATE AXIAL PROFILE OF FRACTIONAL AMMONIA DISSOCIATION

P = 479.5 PSIA
G = 3.12 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

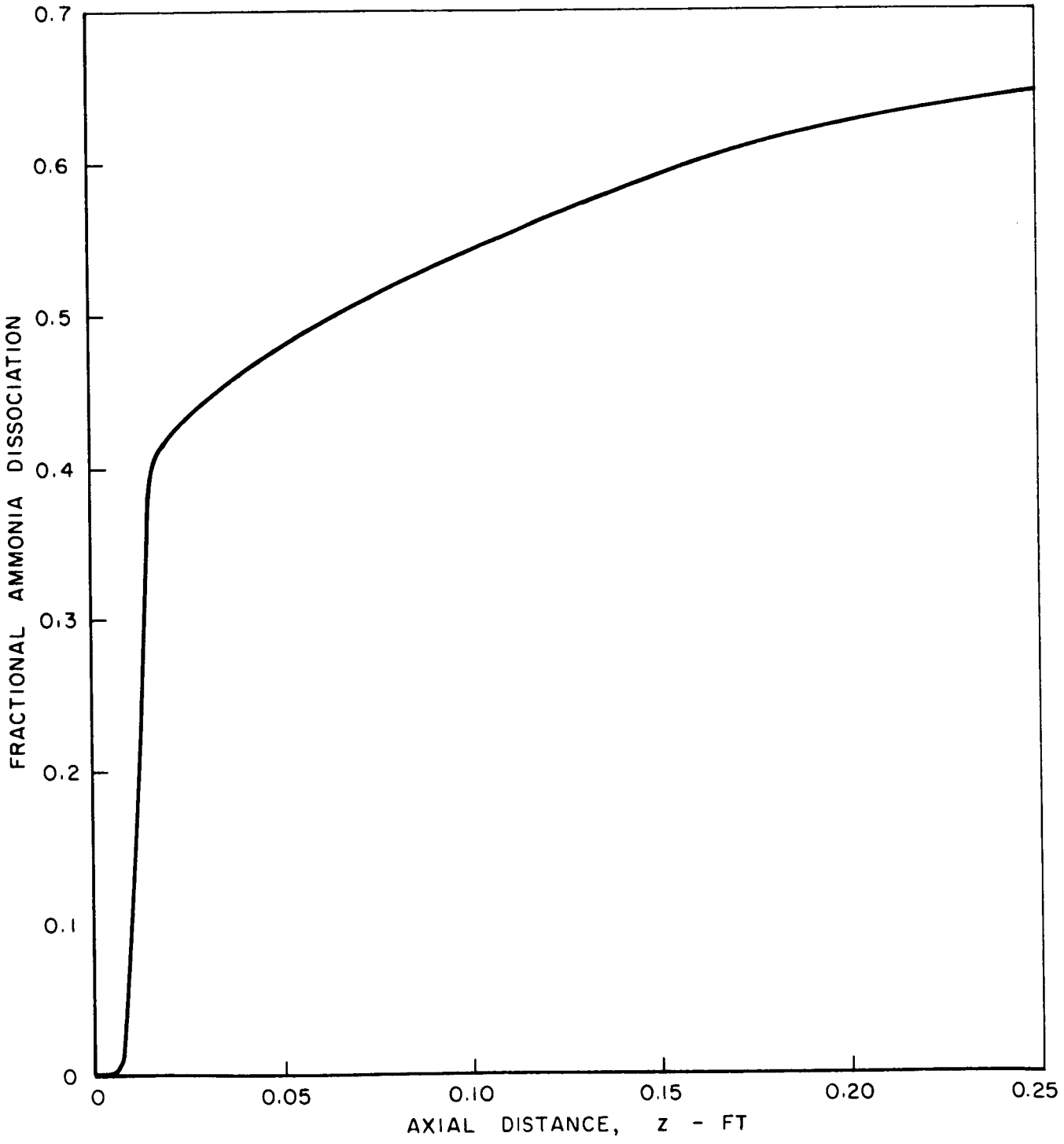
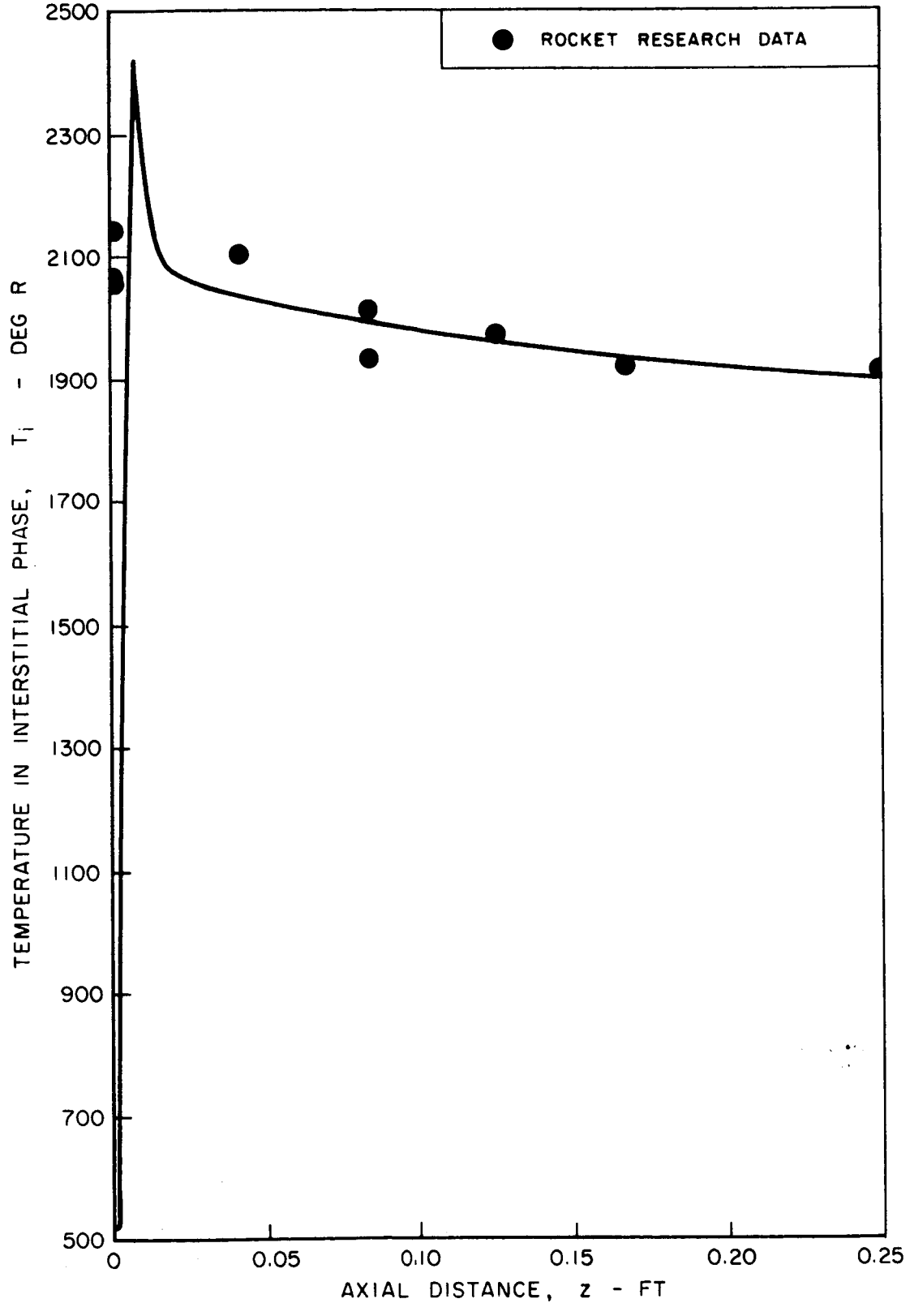


FIG. 8

# STEADY-STATE AXIAL TEMPERATURE PROFILE

P = 1042 PSIA
G = 2.43 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)



# STEADY-STATE AXIAL PROFILES OF MOLE-FRACTIONS OF REACTANTS

P = 1042 PSIA
G = 2.43 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

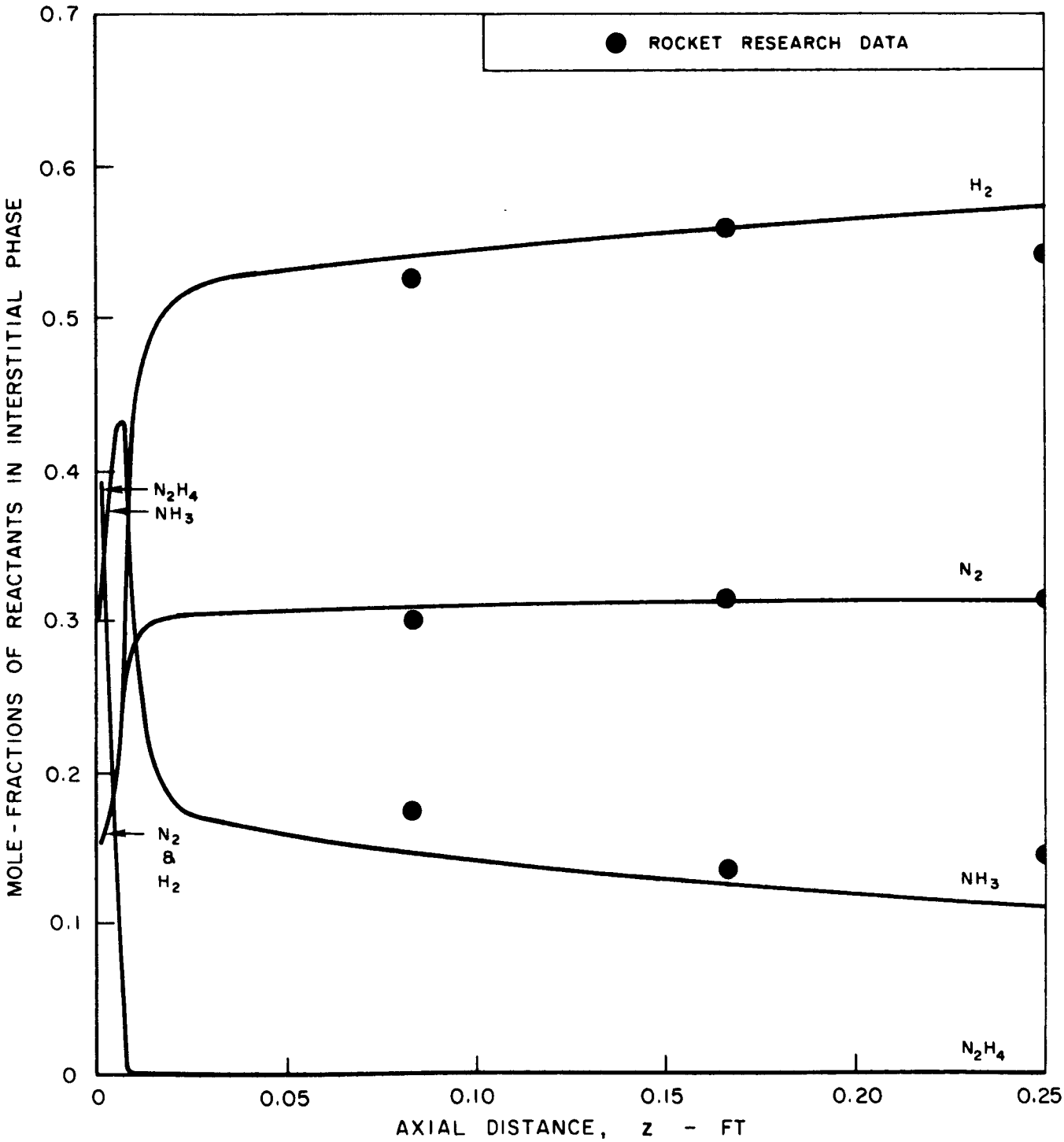


FIG. 10

# STEADY-STATE AXIAL PROFILE OF FRACTIONAL AMMONIA DISSOCIATION

P = 1042 PSIA
G = 2.43 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

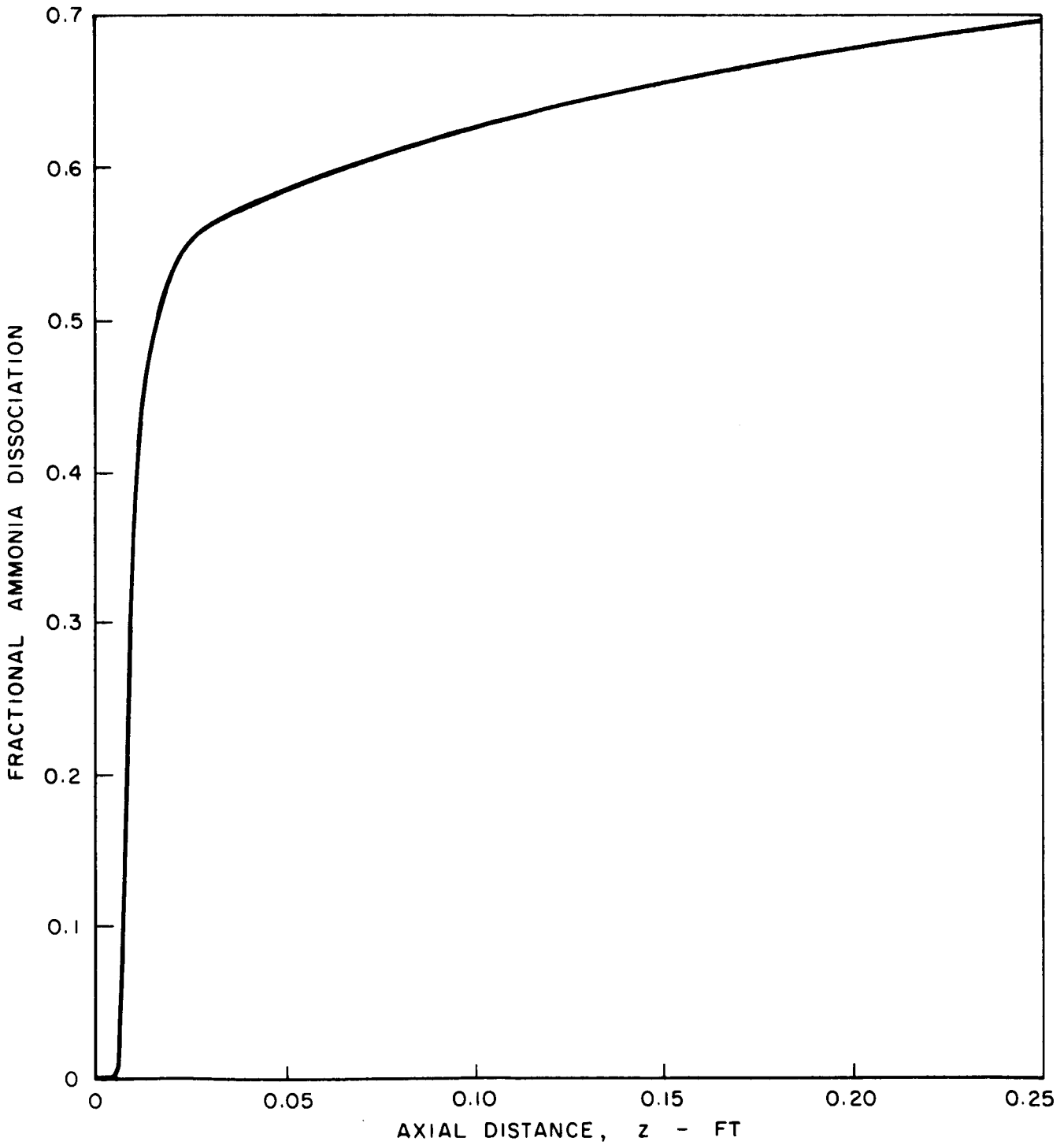


FIG. 11

# STEADY-STATE AXIAL TEMPERATURE PROFILE

P = 974 PSIA
G = 6.29 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

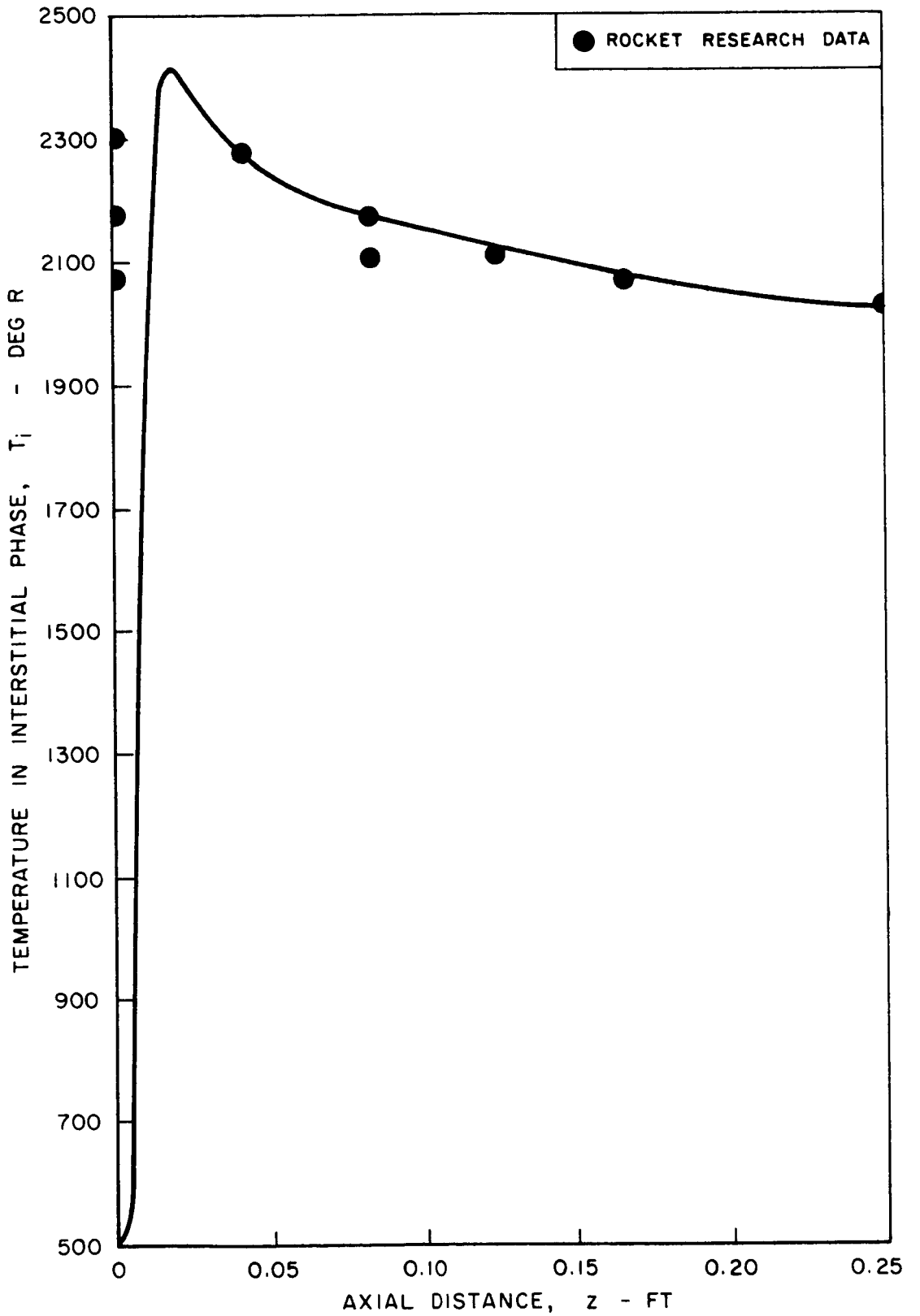


FIG. 12

# STEADY-STATE AXIAL PROFILES OF MOLE-FRACTIONS OF REACTANTS

P = 974 PSIA
G = 6.29 LB/FT$^2$- SEC
STANDARD BED CONFIGURATION (SEE TEXT)

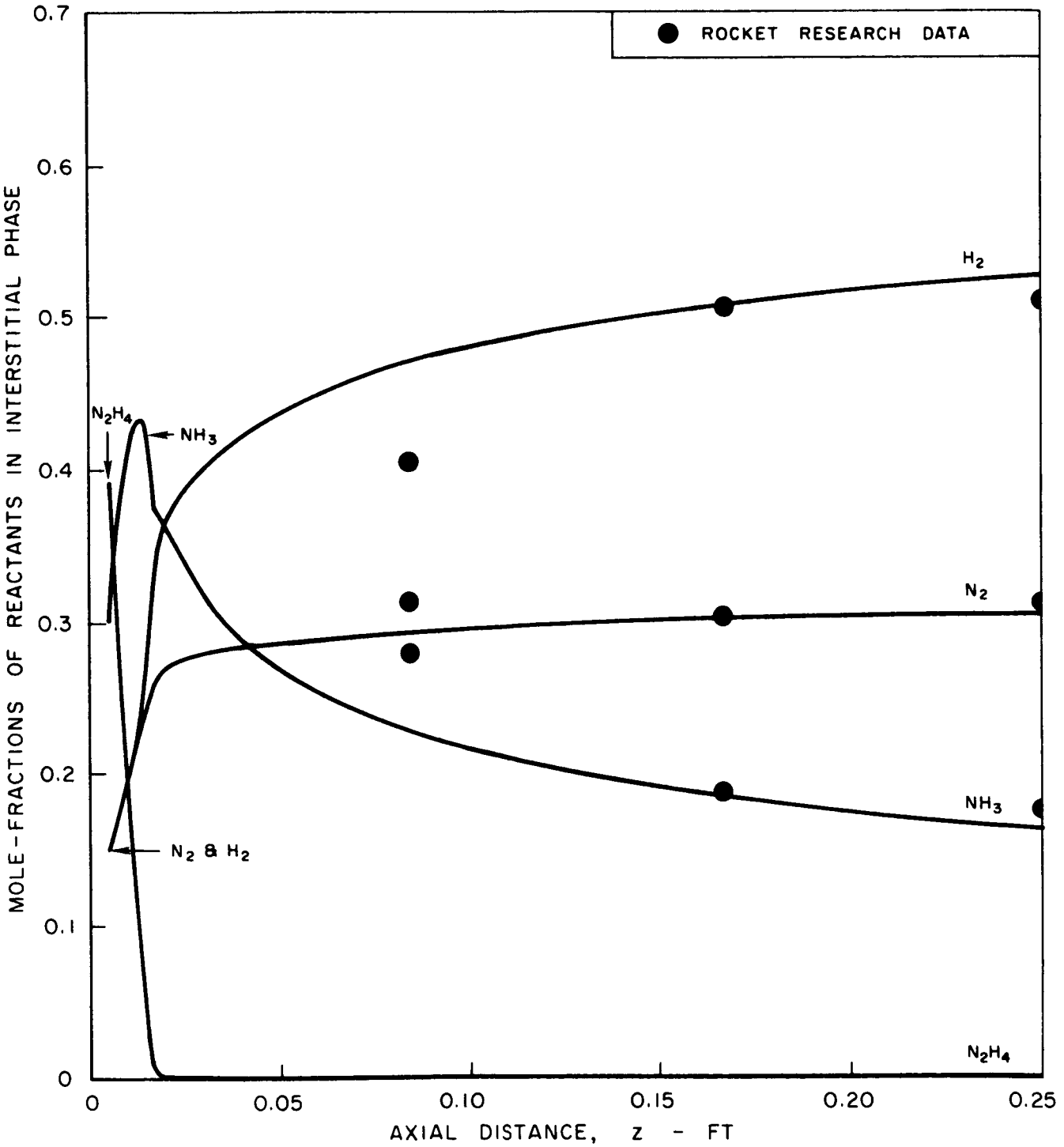


FIG. 13

# STEADY-STATE AXIAL PROFILE OF FRACTIONAL AMMONIA DISSOCIATION

P = 974 PSIA
G = 6.29 LB/FT$^2$-SEC
STANDARD BED CONFIGURATION (SEE TEXT)

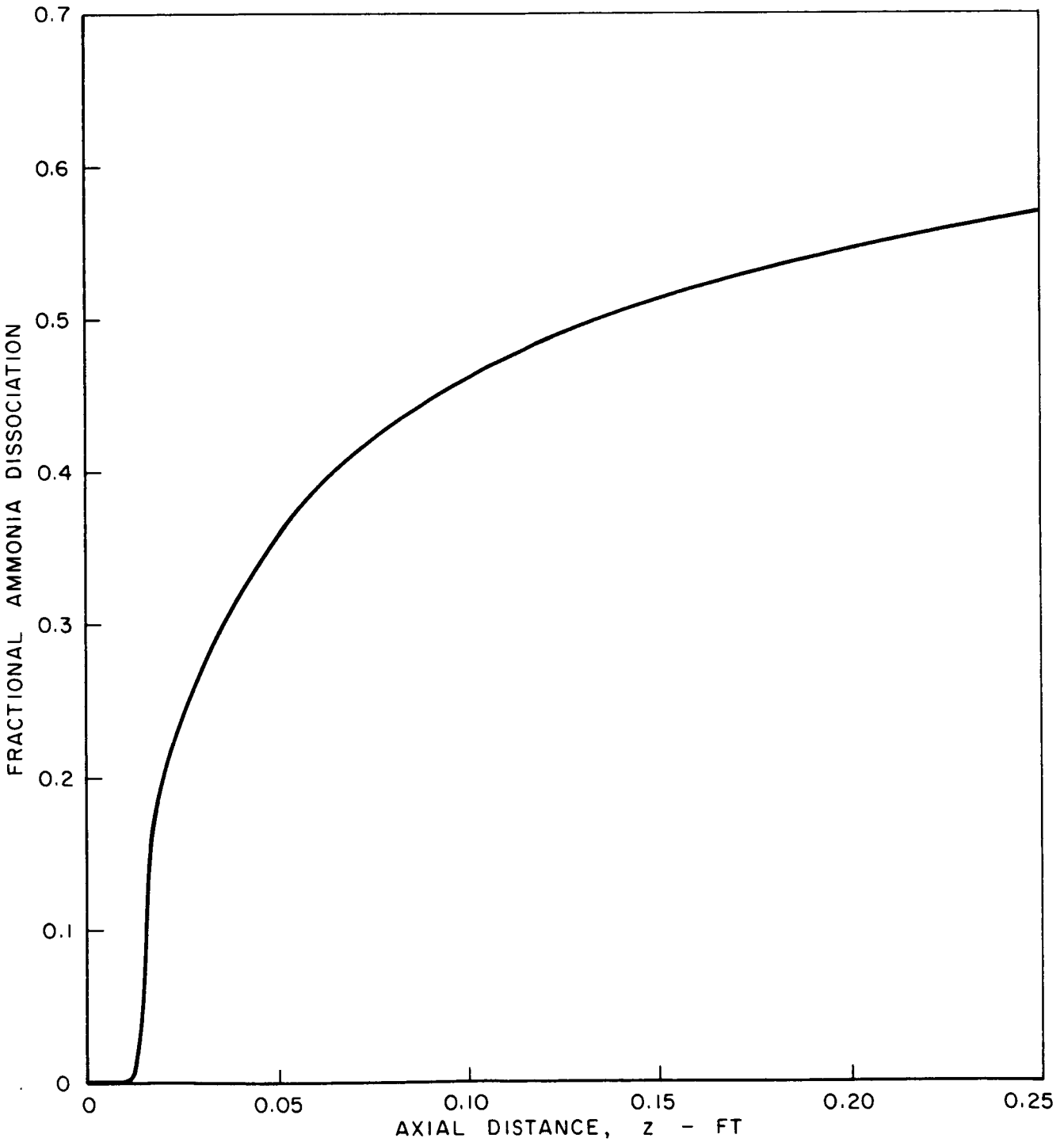
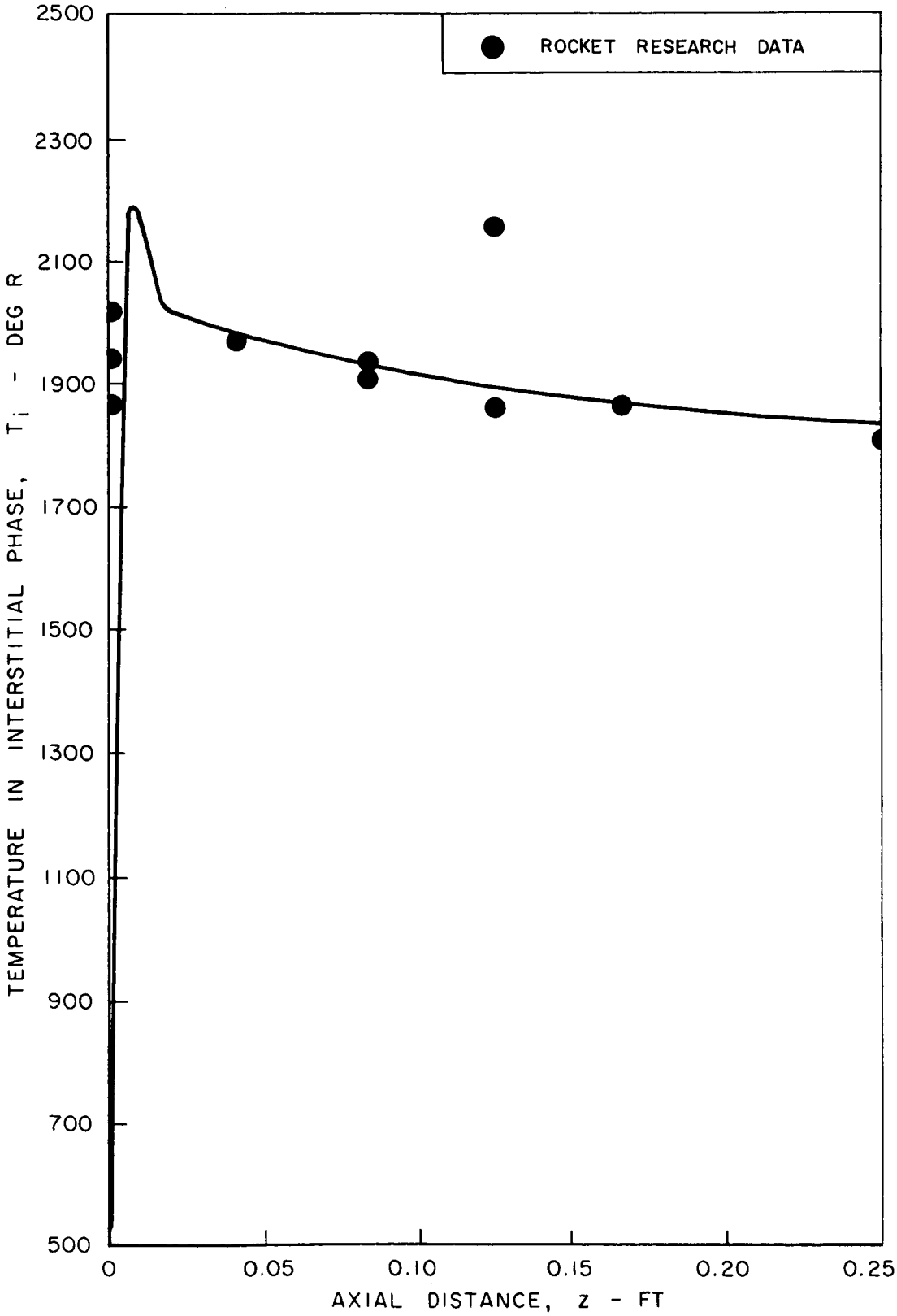


FIG. 14

# STEADY-STATE AXIAL TEMPERATURE PROFILE

P = 217.9 PSIA
G = 1.52 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)



# STEADY-STATE AXIAL PROFILES OF MOLE-FRACTIONS OF REACTANTS

P = 217.9 PSIA
G = 1.52 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

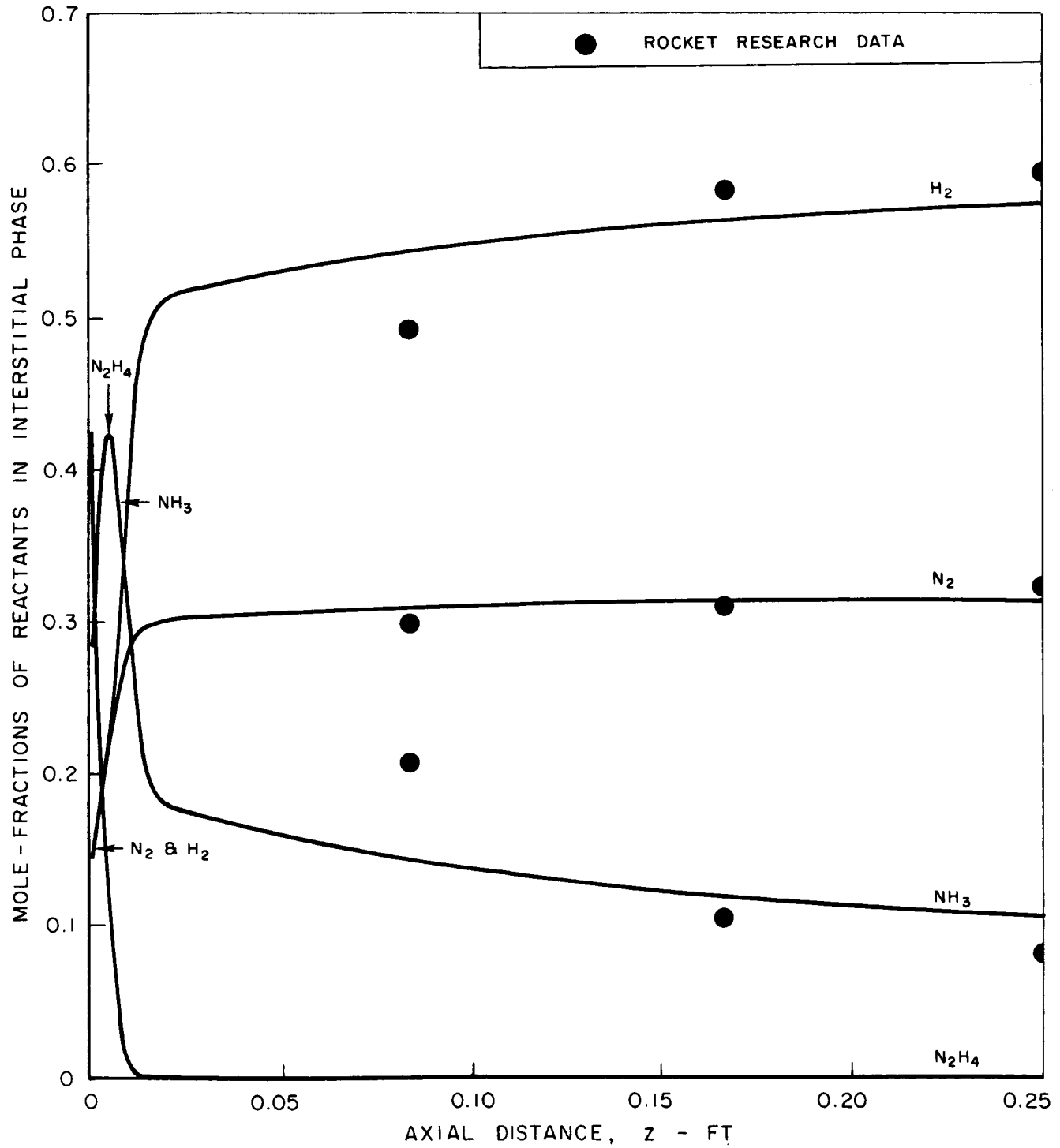


FIG. 16

# STEADY-STATE AXIAL PROFILE OF FRACTIONAL AMMONIA DISSOCIATION

P = 217.9 PSIA
G = 1.52 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

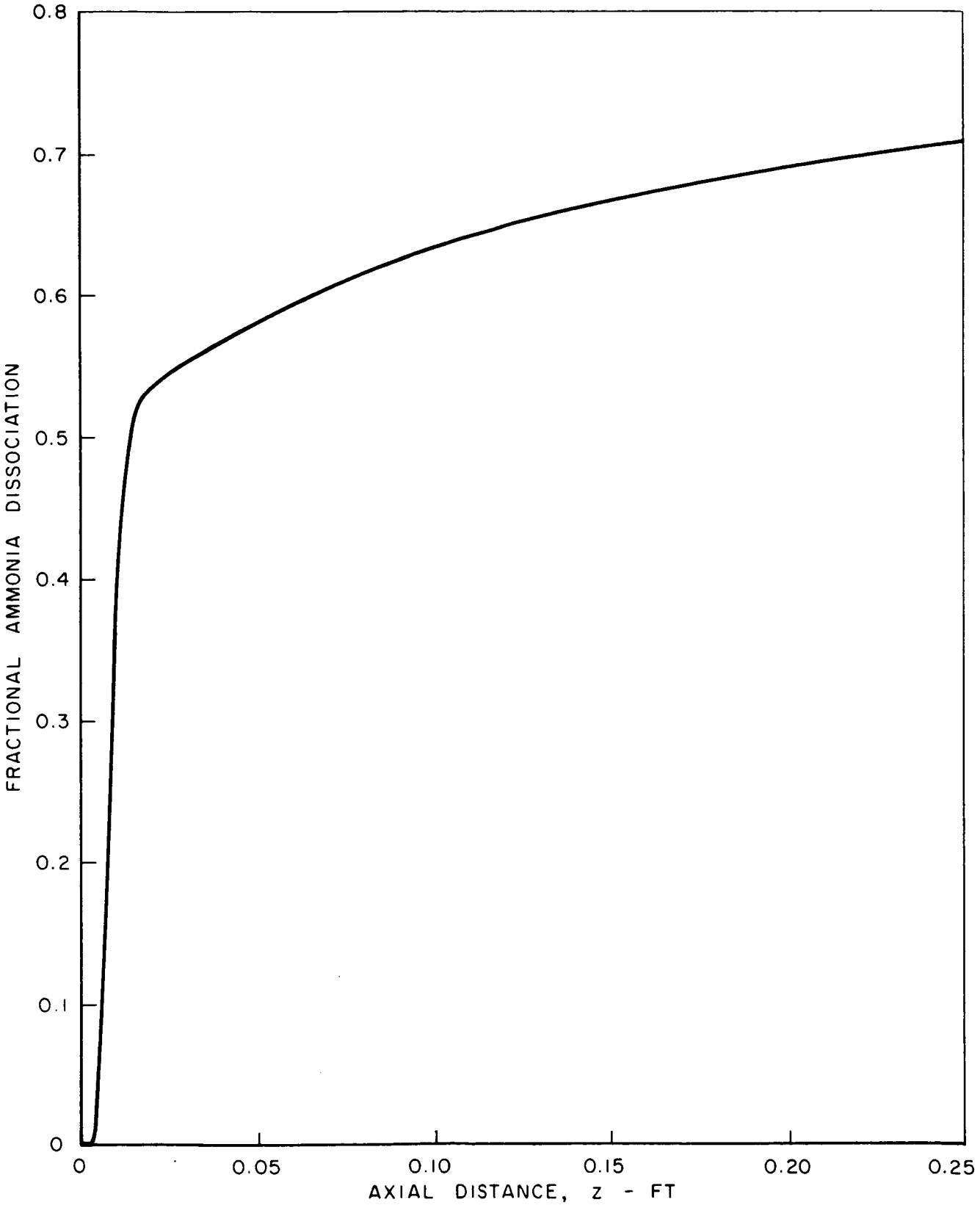


FIG. 17

# STEADY-STATE AXIAL TEMPERATURE PROFILE

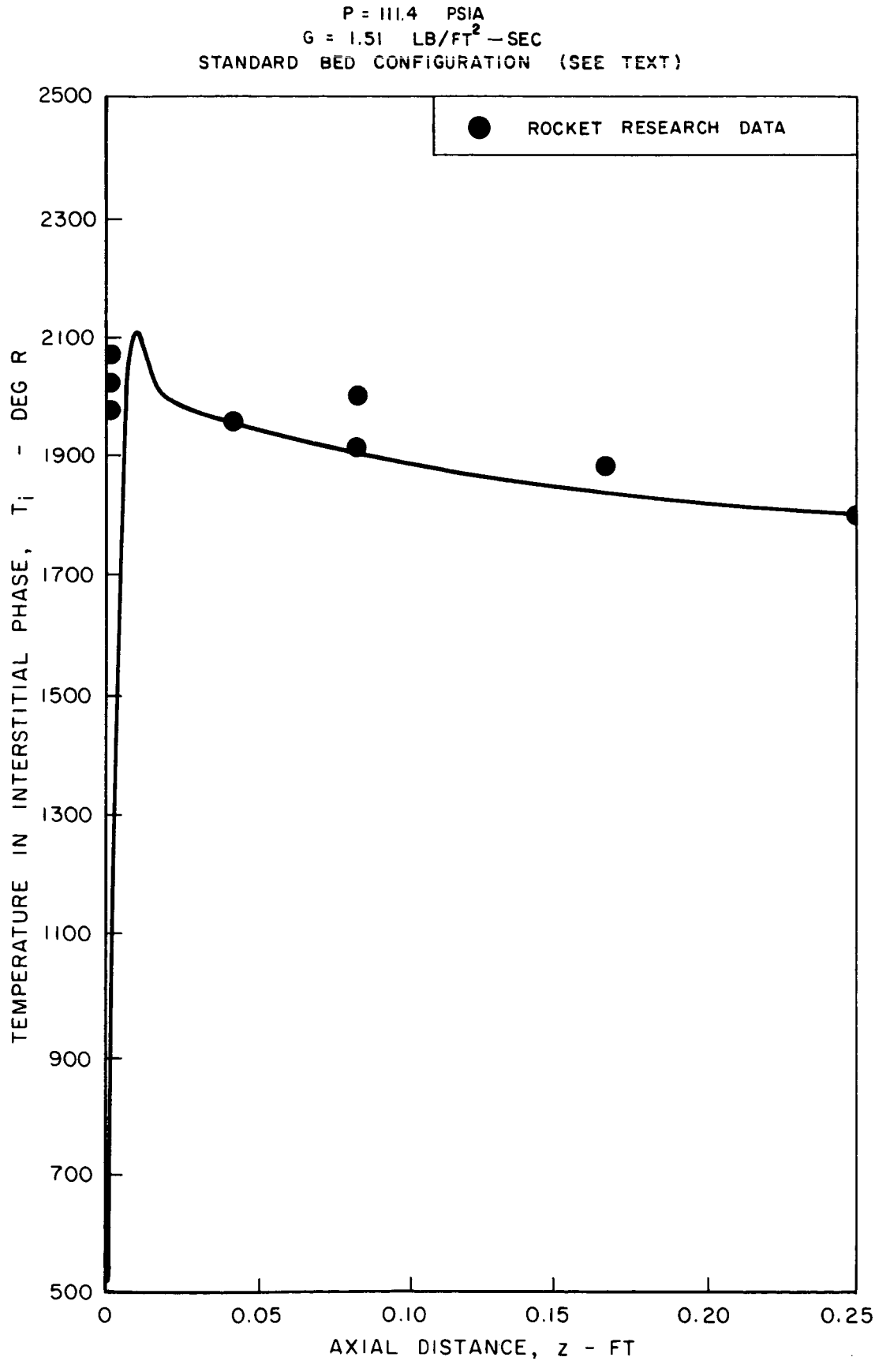


FIG. 18

# STEADY-STATE AXIAL PROFILES OF MOLE-FRACTIONS OF REACTANTS

P = 111.4 PSIA
G = 1.51 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

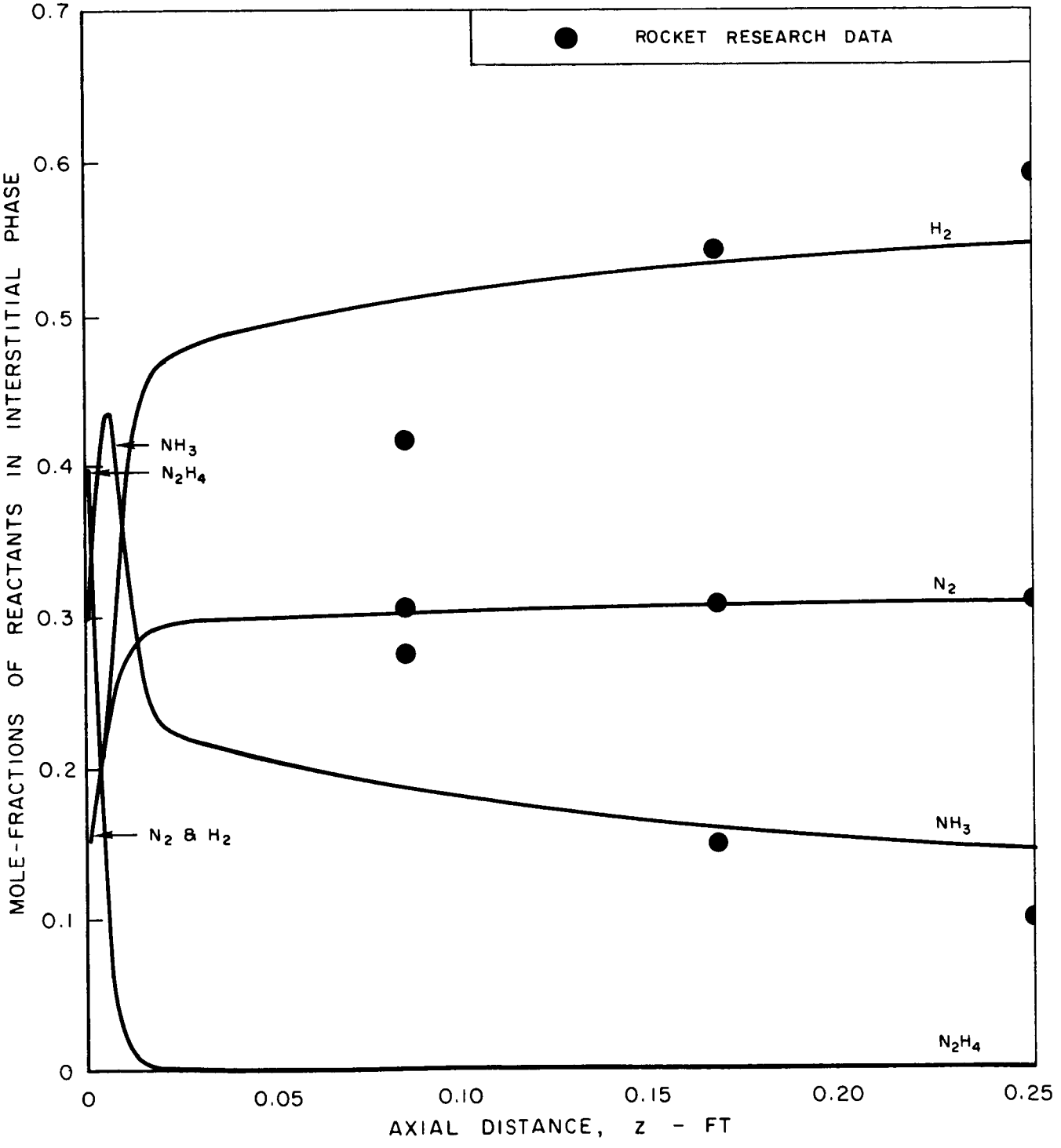


FIG. 19

# STEADY-STATE AXIAL PROFILE OF FRACTIONAL AMMONIA DISSOCIATION

P = 111.4 PSIA
G = 1.51 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

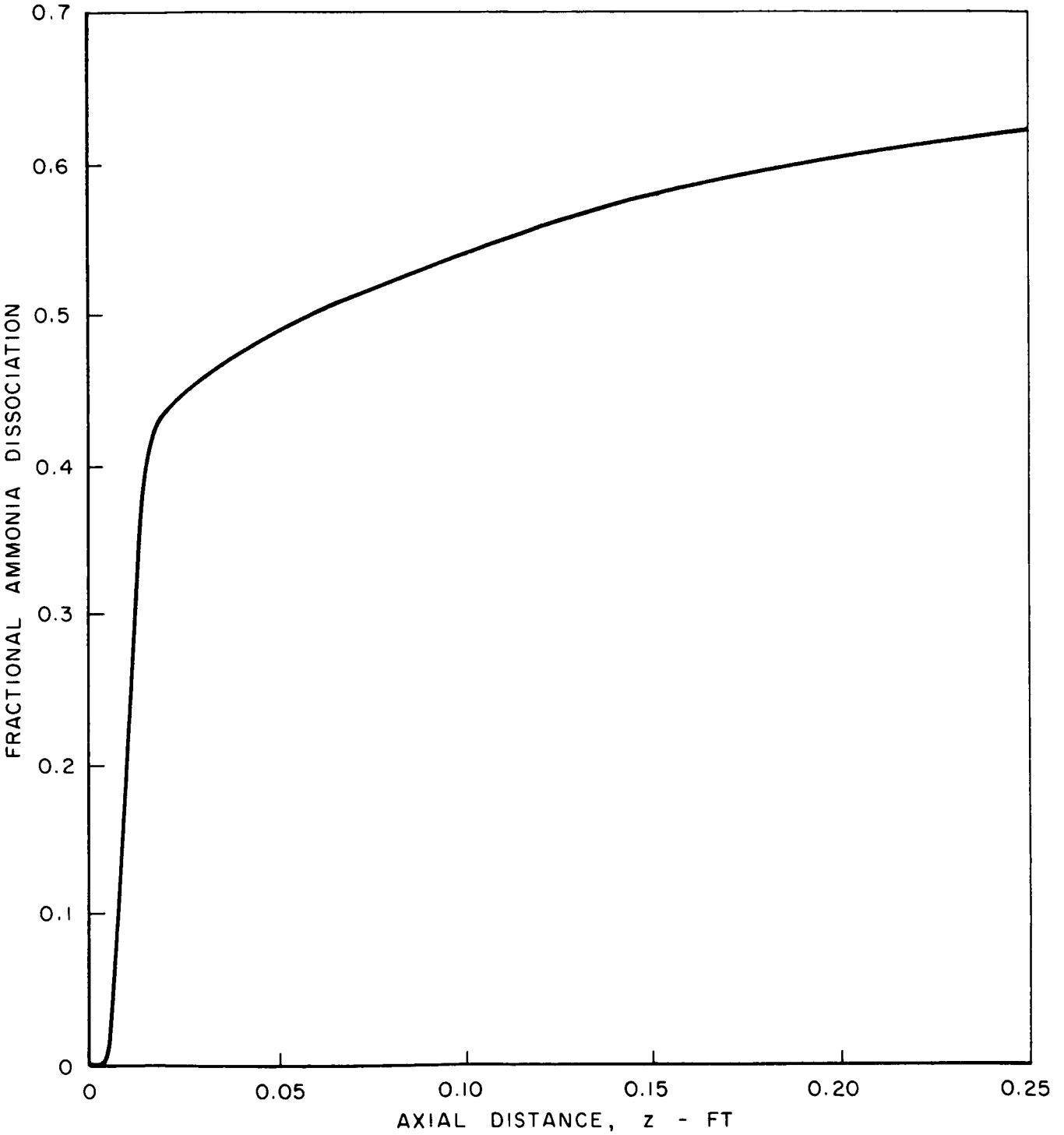


FIG. 20

# STEADY-STATE AXIAL TEMPERATURE PROFILES FOR VARIOUS CHAMBER PRESSURES

G = 3.0 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

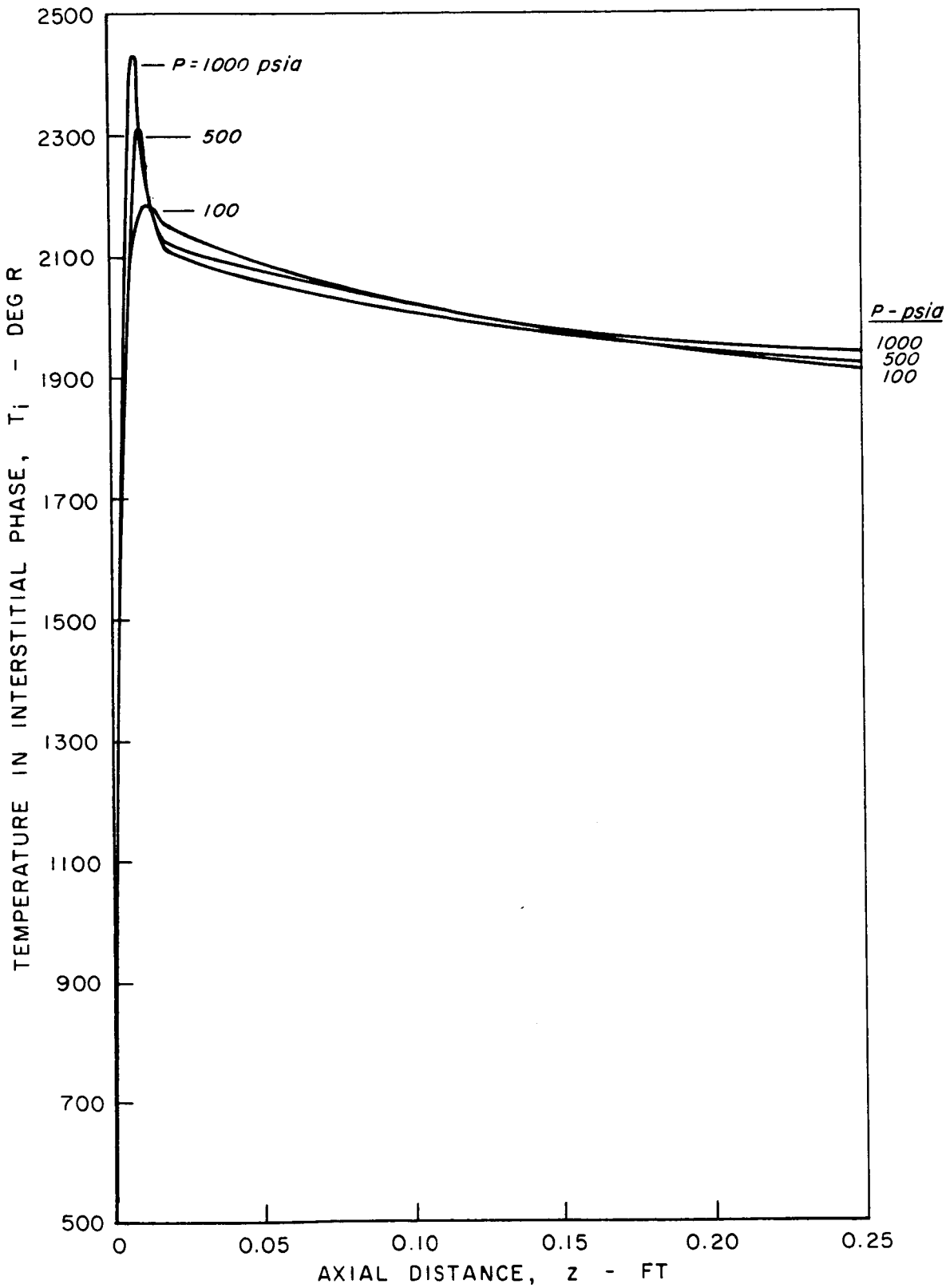


FIG. 21

# STEADY-STATE AXIAL TEMPERATURE PROFILES FOR VARIOUS MASS FLOW RATES

P = 100 PSIA

STANDARD BED CONFIGURATION (SEE TEXT)

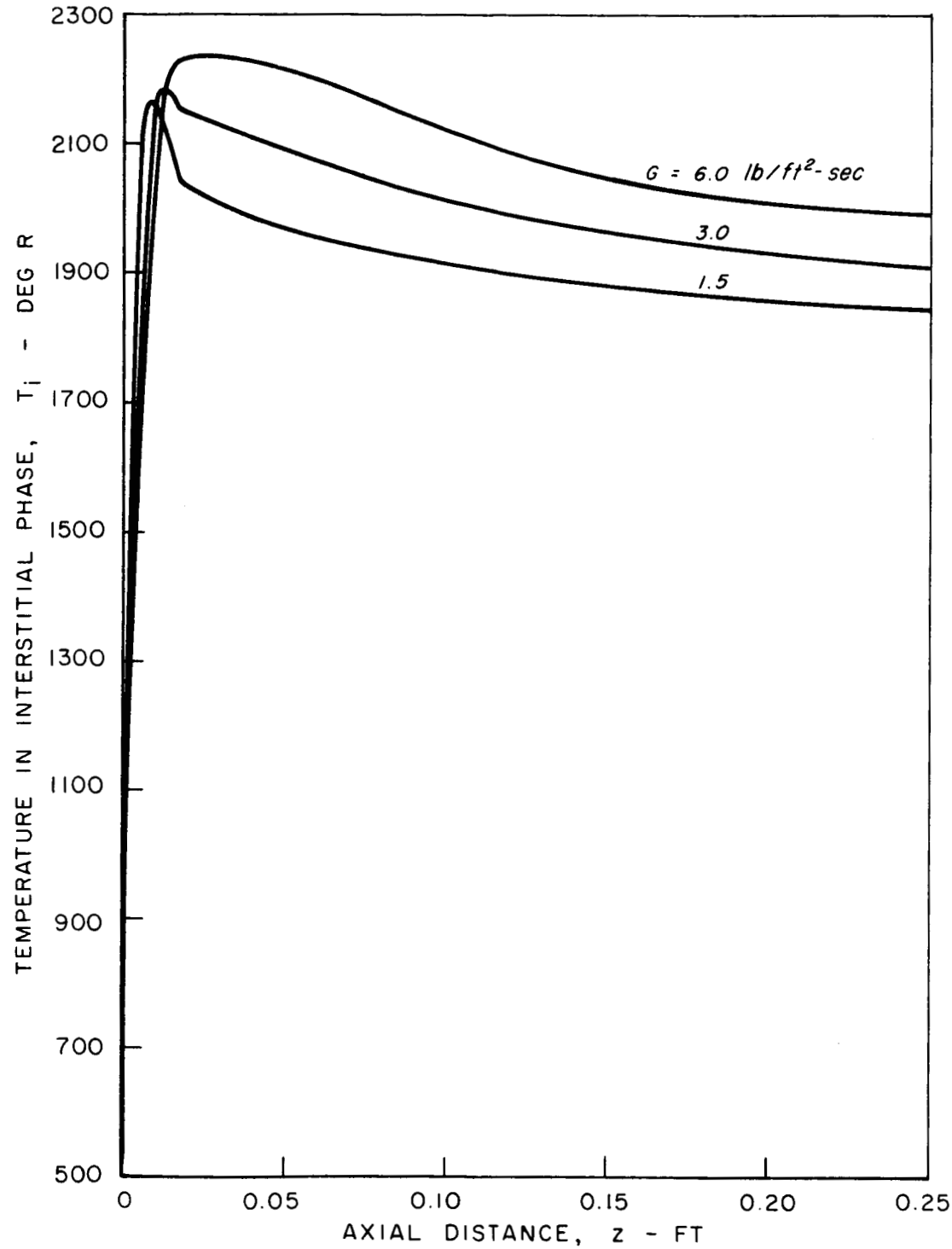
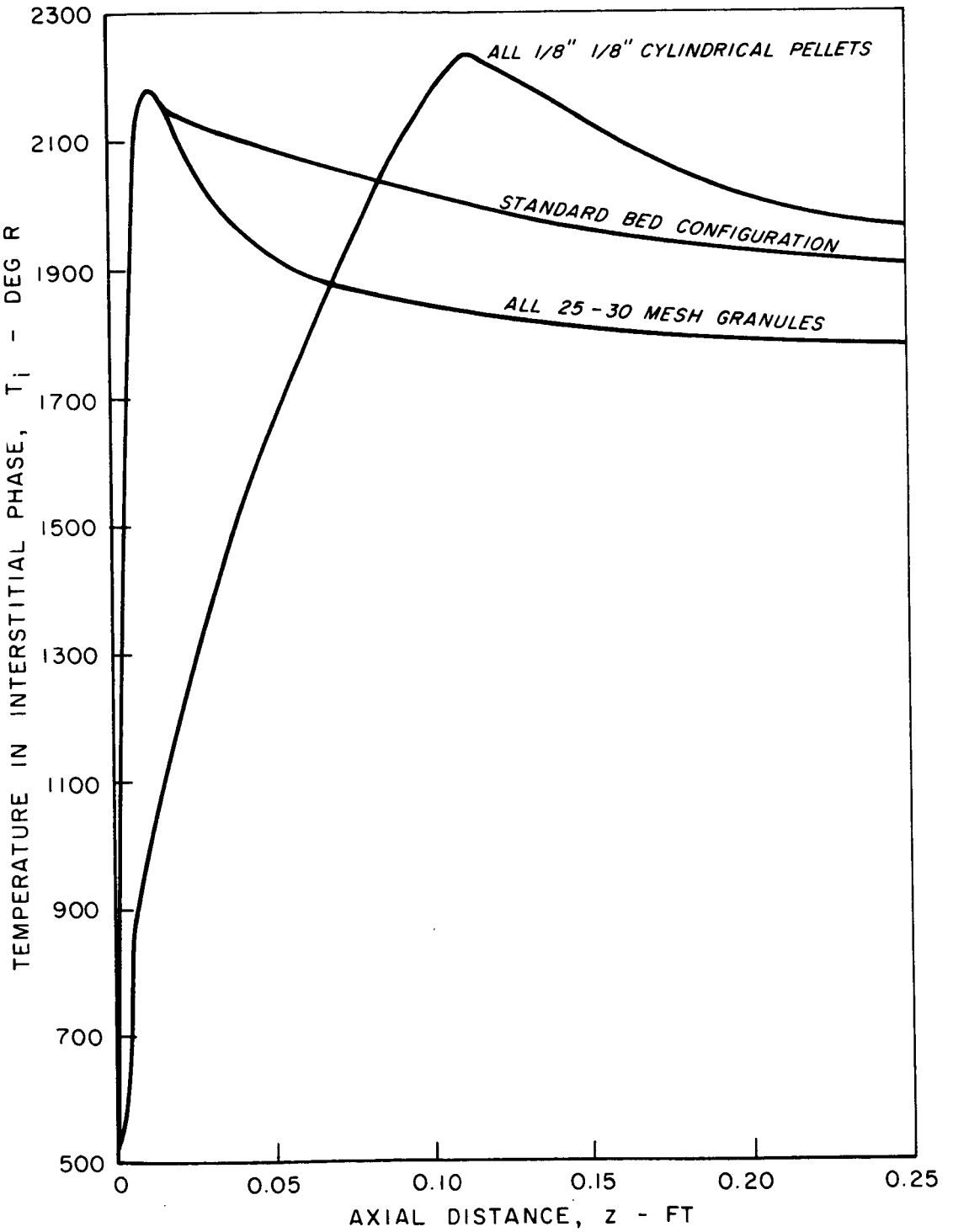


FIG. 22

# STEADY-STATE AXIAL TEMPERATURE PROFILES FOR VARIOUS CATALYTIC BED CONFIGURATIONS

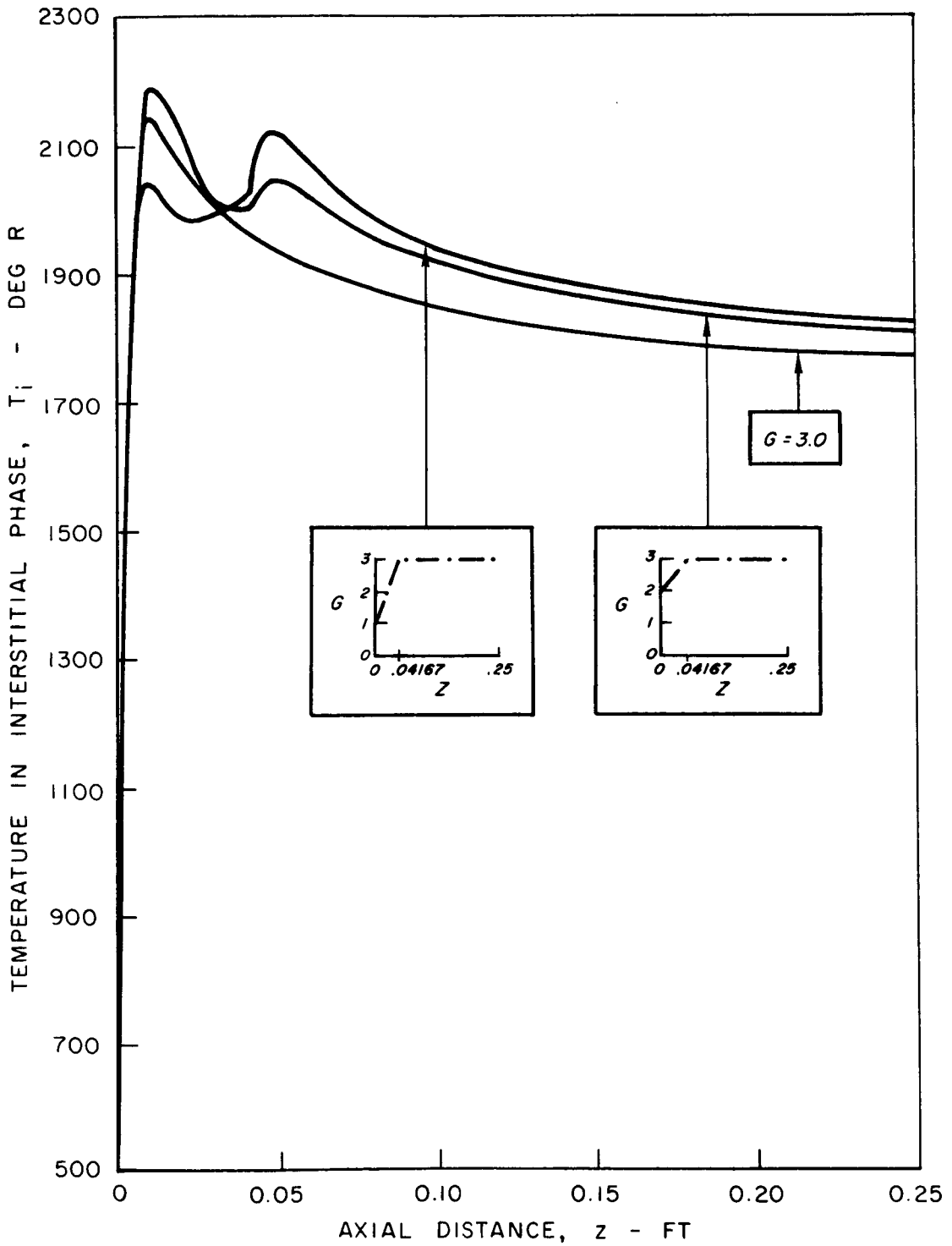
P = 100 PSIA

G = 3.0 LB/FT$^2$-SEC



# STEADY-STATE AXIAL TEMPERATURE PROFILES FOR VARIOUS HYDRAZINE AXIAL INJECTION PROFILES

P = 100 PSIA
BED CONFIGURATION: ALL 25-30 MESH GRANULES



# COMPARISON OF STEADY-STATE AXIAL TEMPERATURE PROFILES COMPUTED FROM STEADY-STATE AND TRANSIENT MODELS

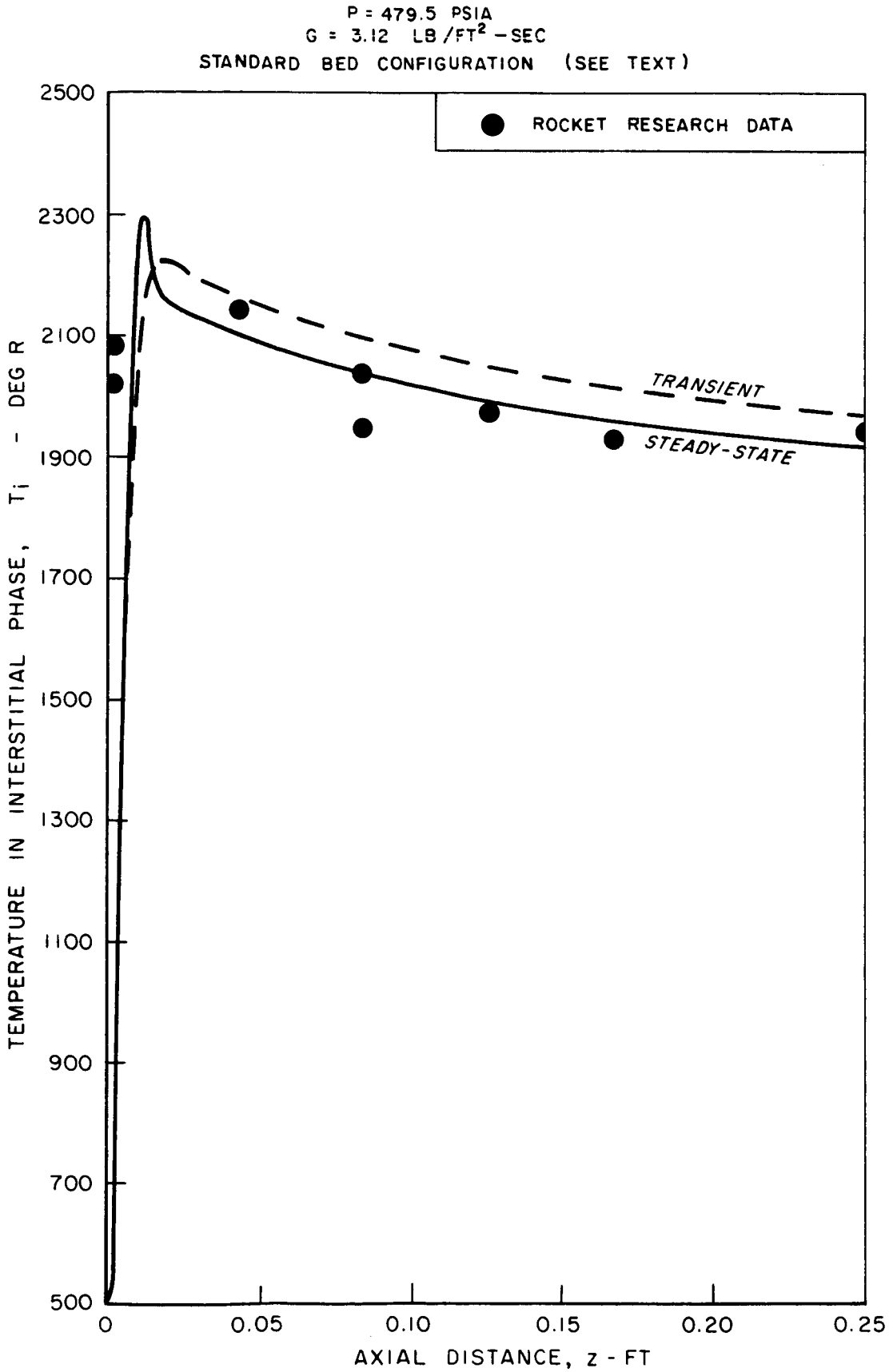


FIG. 25

# COMPARISON OF STEADY-STATE AXIAL PROFILES OF MOLE-FRACTION OF AMMONIA COMPUTED FROM STEADY-STATE AND TRANSIENT MODELS

P = 479.5 PSIA
G = 3.12 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

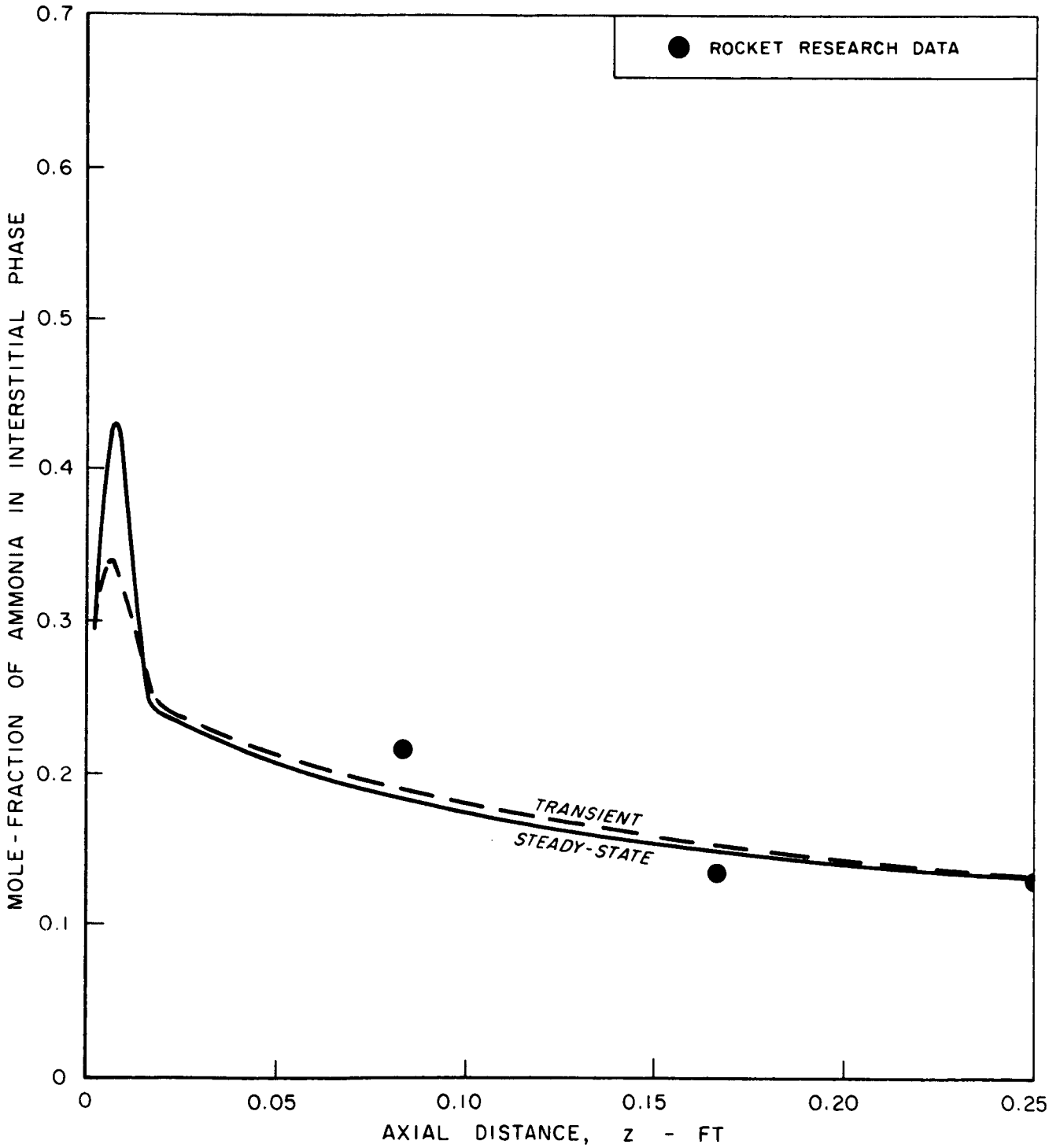


FIG. 26

# TRANSIENT AXIAL TEMPERATURE PROFILES

P (FEED PRESSURE) = 479.5 PSIA
G = 3.12 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

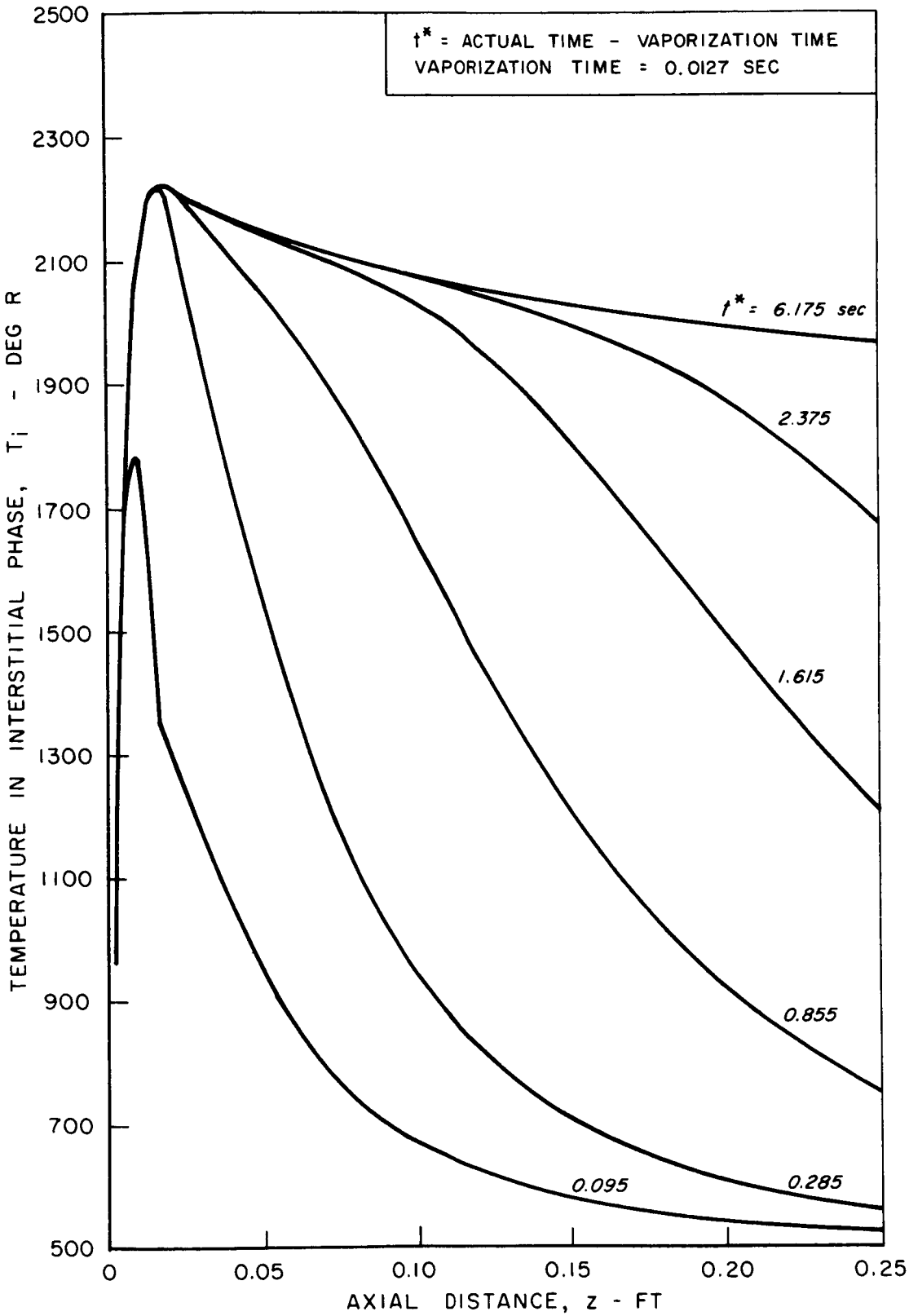


FIG. 27

# VARIATION OF TEMPERATURE WITH TIME AT SEVERAL AXIAL POSITIONS

P (FEED PRESSURE) = 479.5 PSIA  G = 3.12 LB/FT$^2$-SEC  STANDARD BED CONFIGURATION (SEE TEXT)

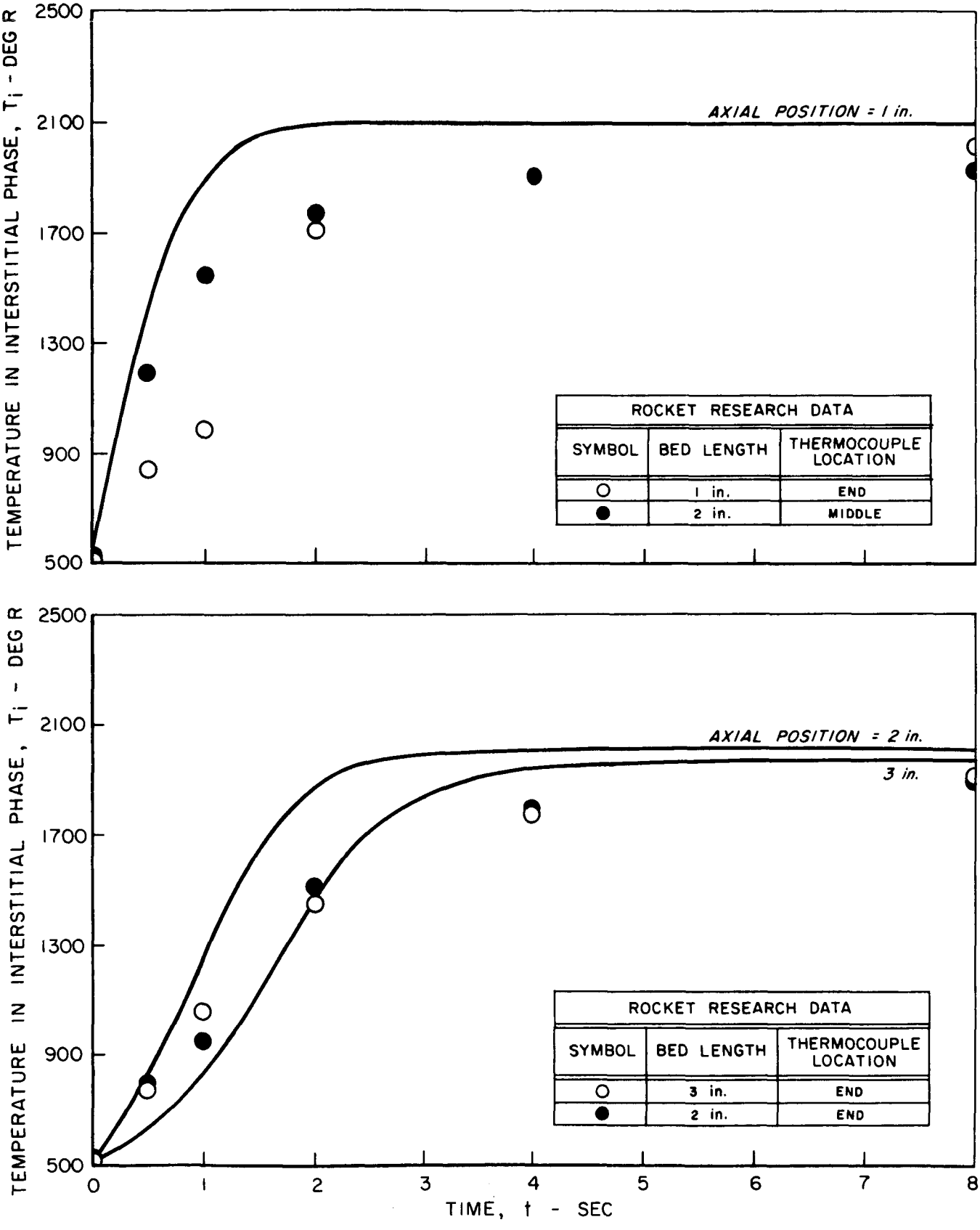


FIG. 28

# TRANSIENT AXIAL PROFILES OF MOLE-FRACTION OF HYDRAZINE

P (FEED PRESSURE) = 479.5 PSIA
G = 3.12 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

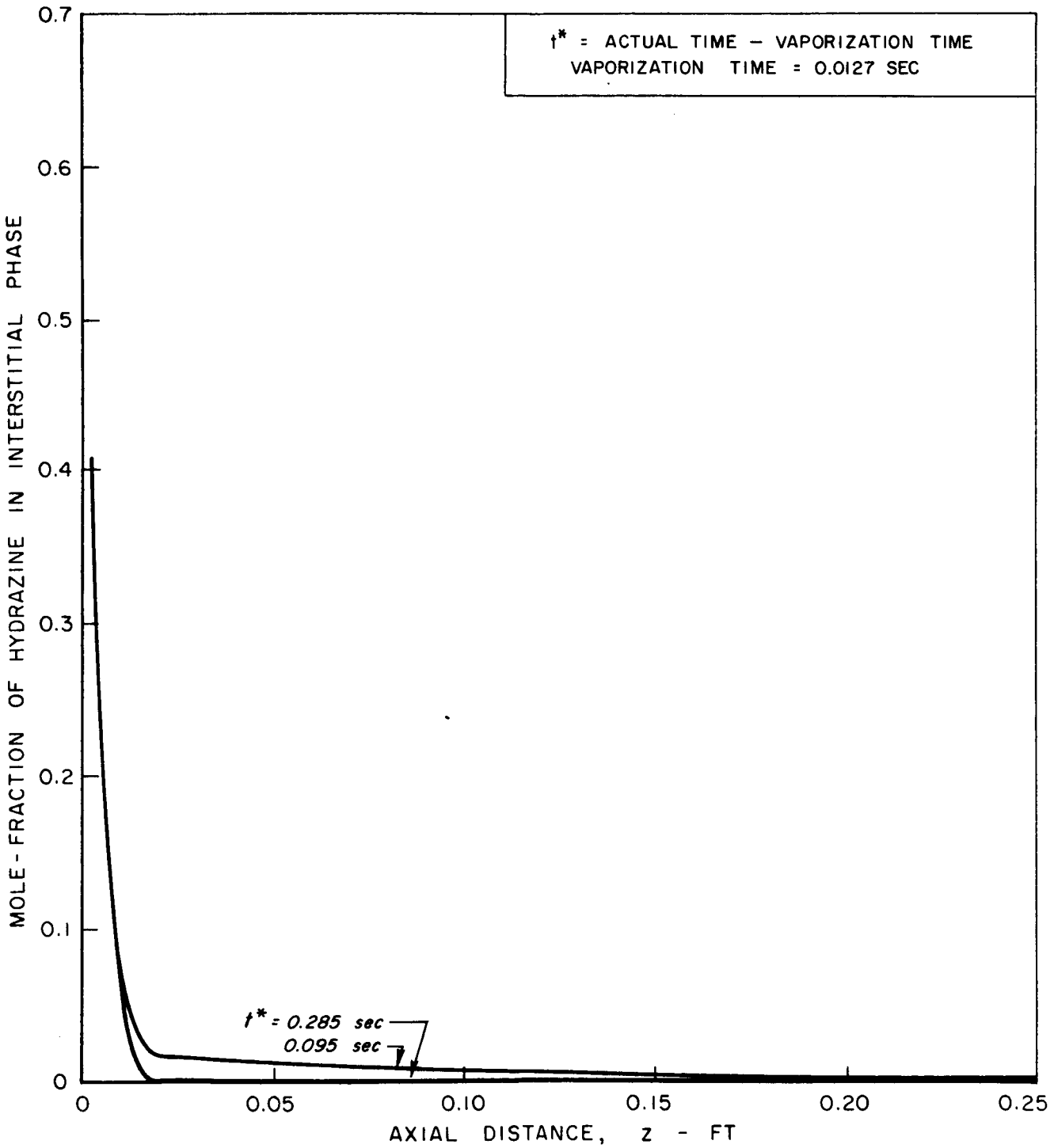


FIG. 29

# TRANSIENT AXIAL PROFILES OF MOLE-FRACTION OF AMMONIA

P(FEED PRESSURE) = 479.5 PSIA
G = 3.12 LB/FT$^2$-SEC
STANDARD BED CONFIGURATION (SEE TEXT)

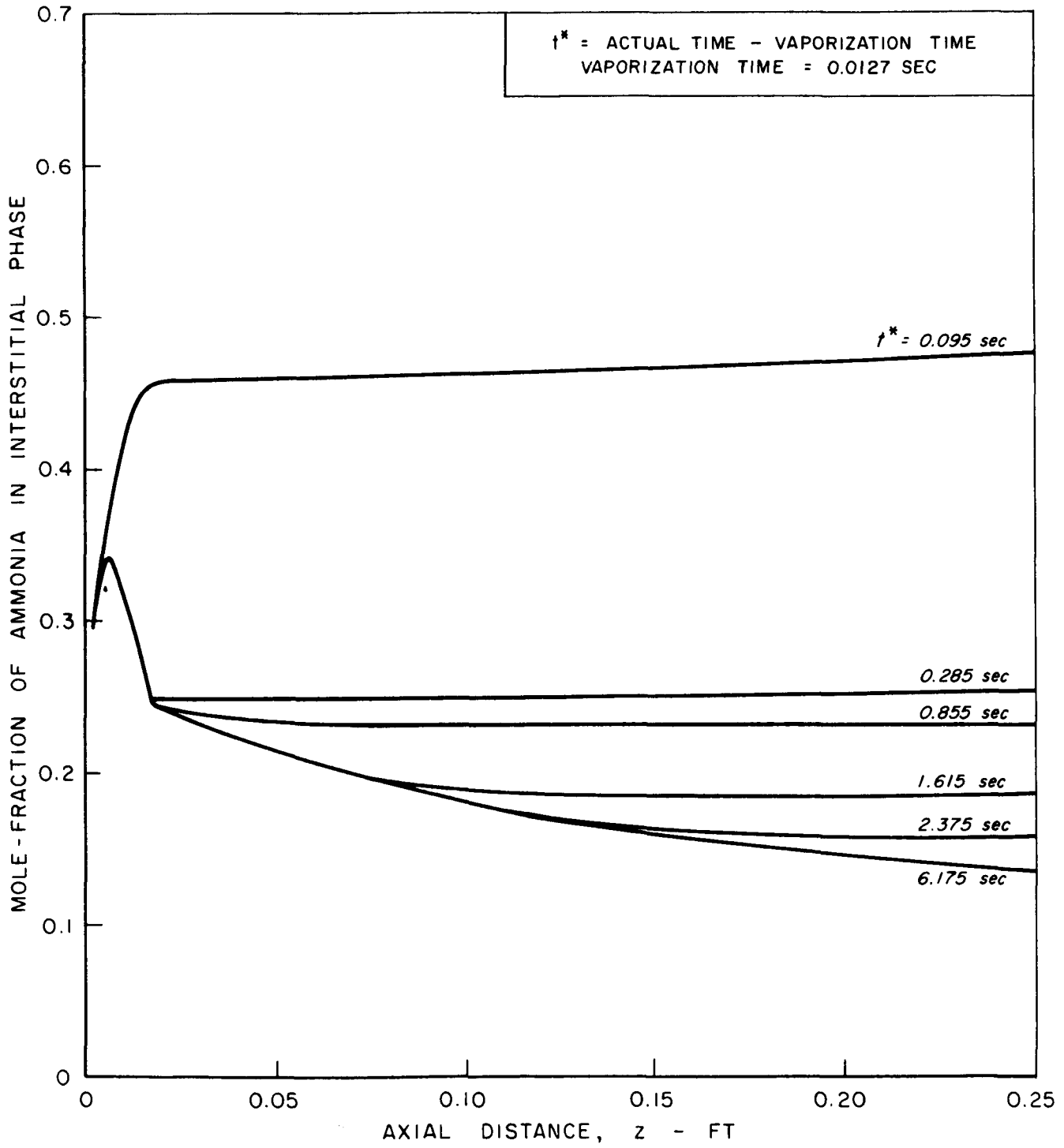


FIG. 30

# TRANSIENT AXIAL PROFILES OF MOLE-FRACTION OF NITROGEN

P (FEED PRESSURE) = 479.5 PSIA
G = 3.12 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

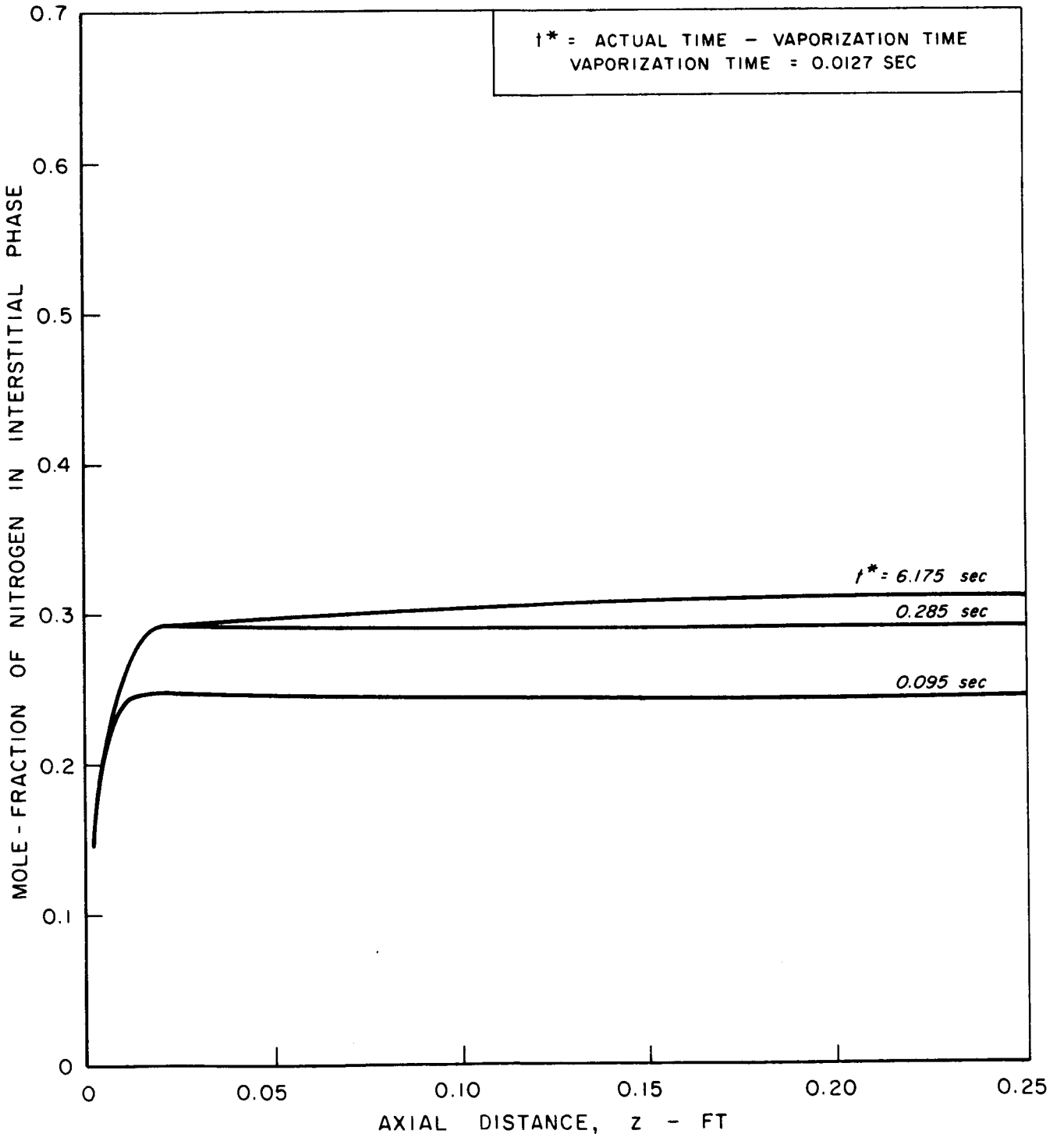


FIG. 31

# TRANSIENT AXIAL PROFILES OF MOLE-FRACTION OF HYDROGEN

P (FEED PRESSURE) = 479.5 PSIA
G = 3.12 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

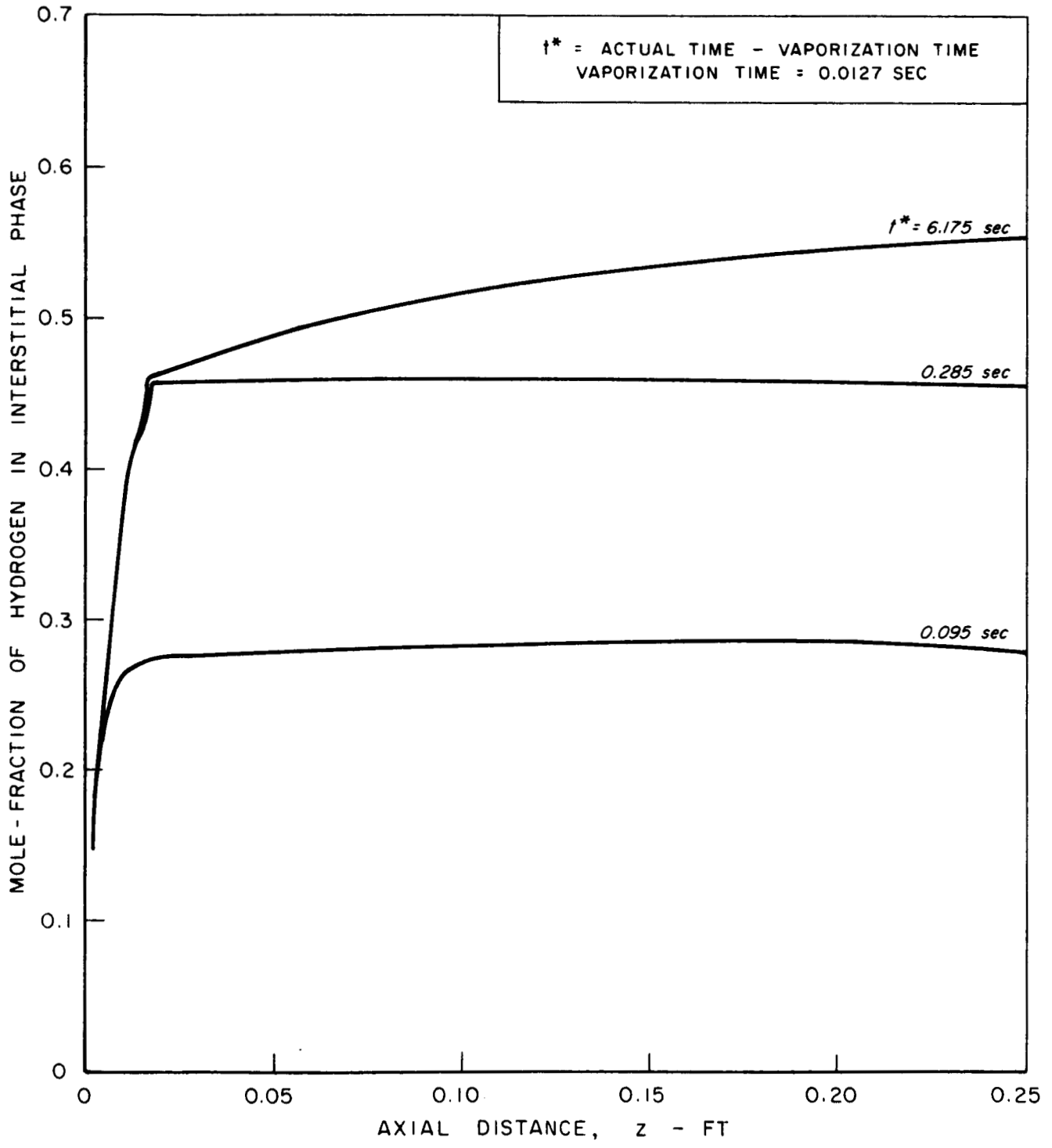


FIG. 32

# VARIATION OF CHAMBER EXIT PRESSURE WITH TIME

P (FEED PRESSURE) = 479.5 PSIA
G = 3.12 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

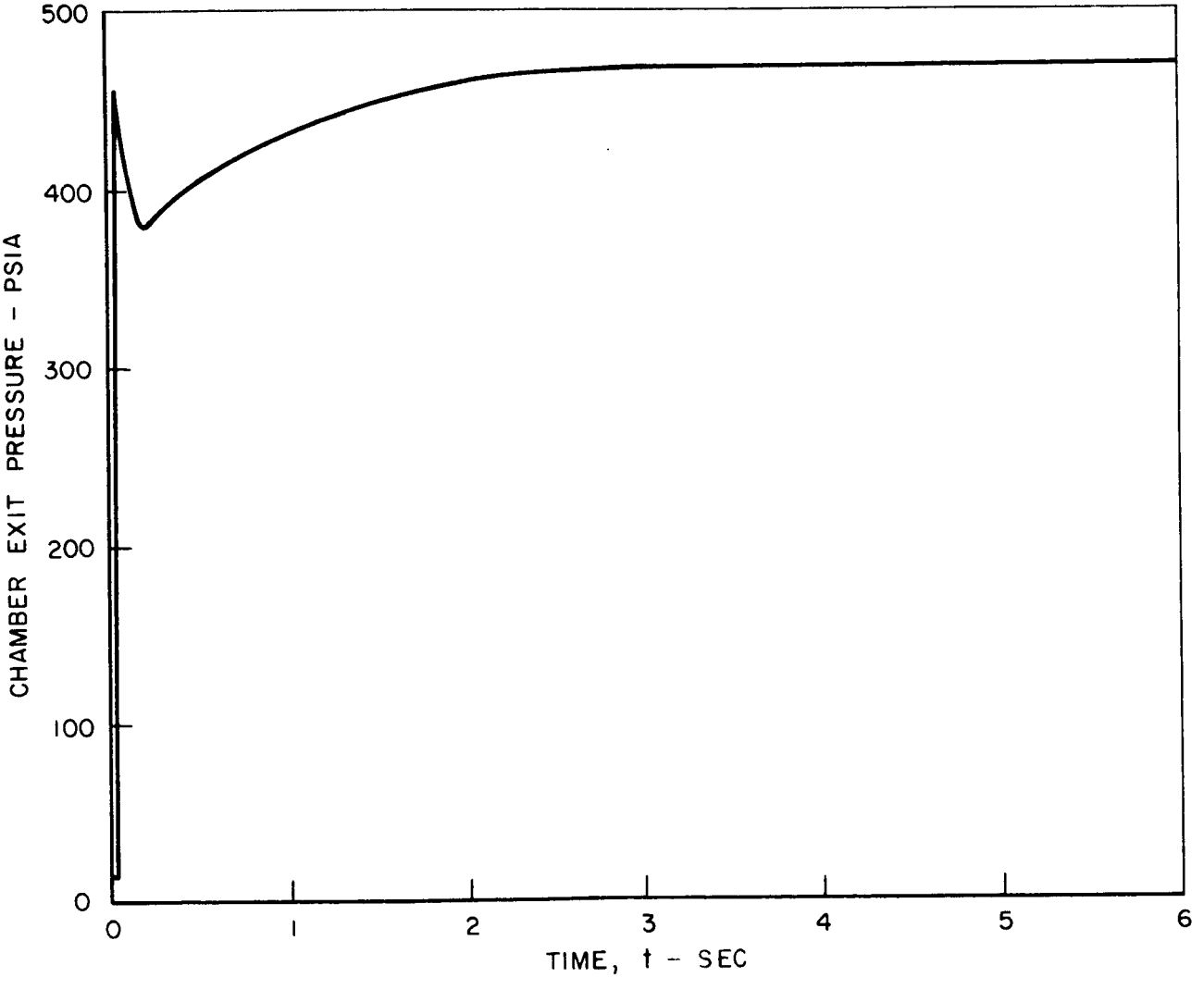


FIG. 33

# TRANSIENT AXIAL TEMPERATURE PROFILES

P (FEED PRESSURE) = 1042 PSIA
G = 2.43 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

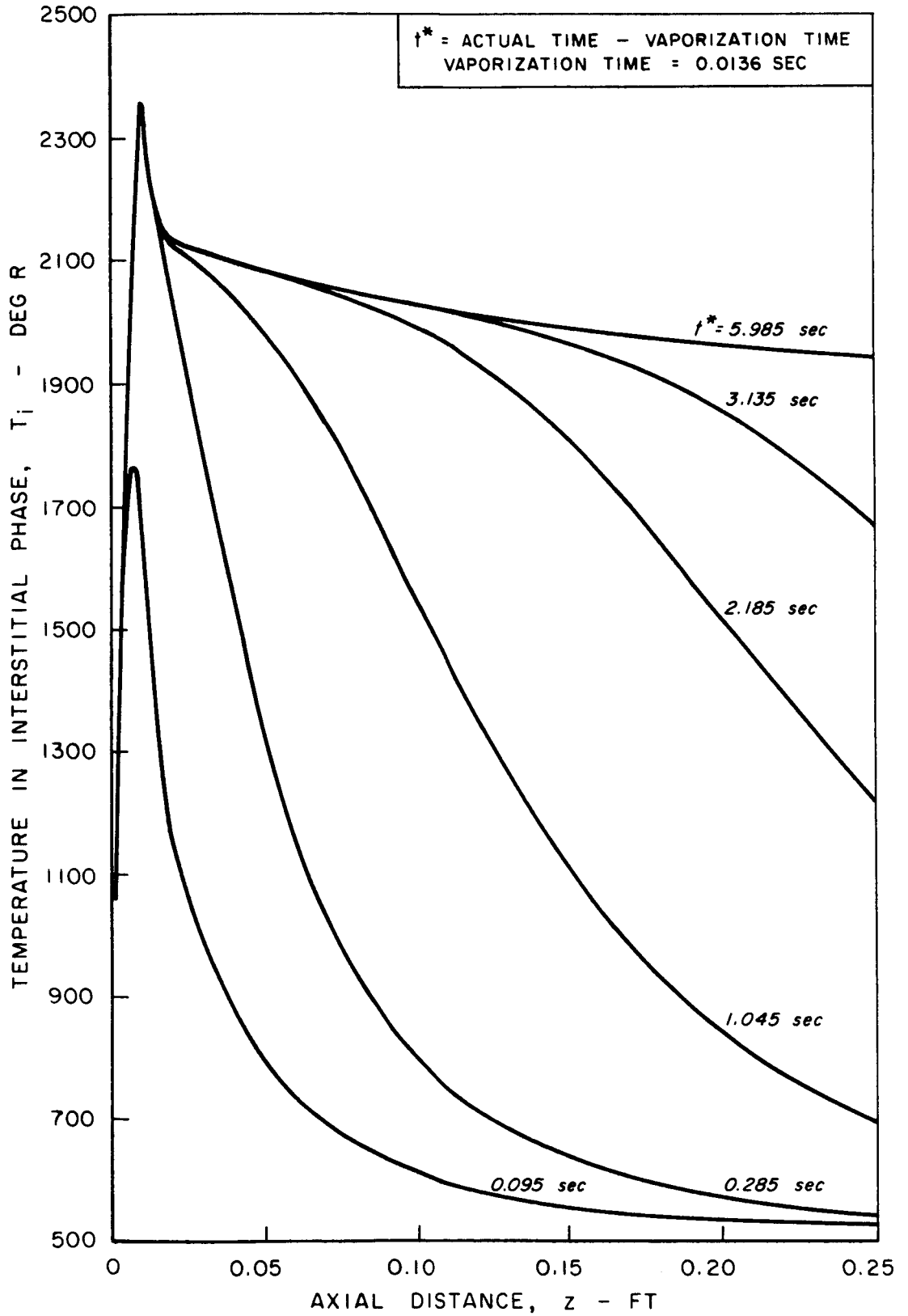


FIG. 34

# VARIATION OF TEMPERATURE WITH TIME AT SEVERAL AXIAL POSITIONS

P(FEED PRESSURE) = 1042 PSIA  G = 2.43 LB/FT$^2$ - SEC  STANDARD BED CONFIGURATION (SEE TEXT)

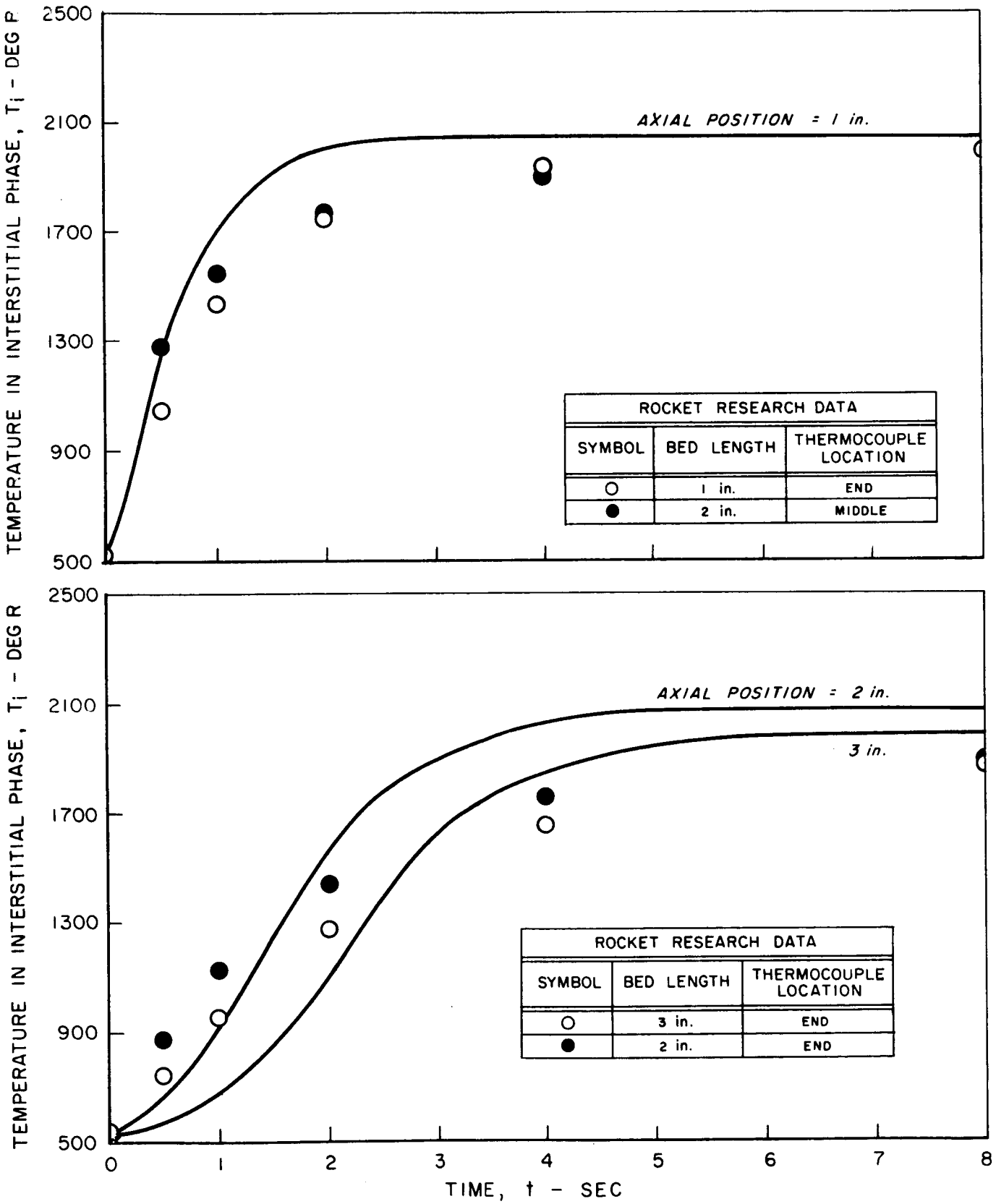


FIG. 35

# VARIATION OF CHAMBER EXIT PRESSURE WITH TIME

P (FEED PRESSURE) = 1042 PSIA
G = 2.43 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

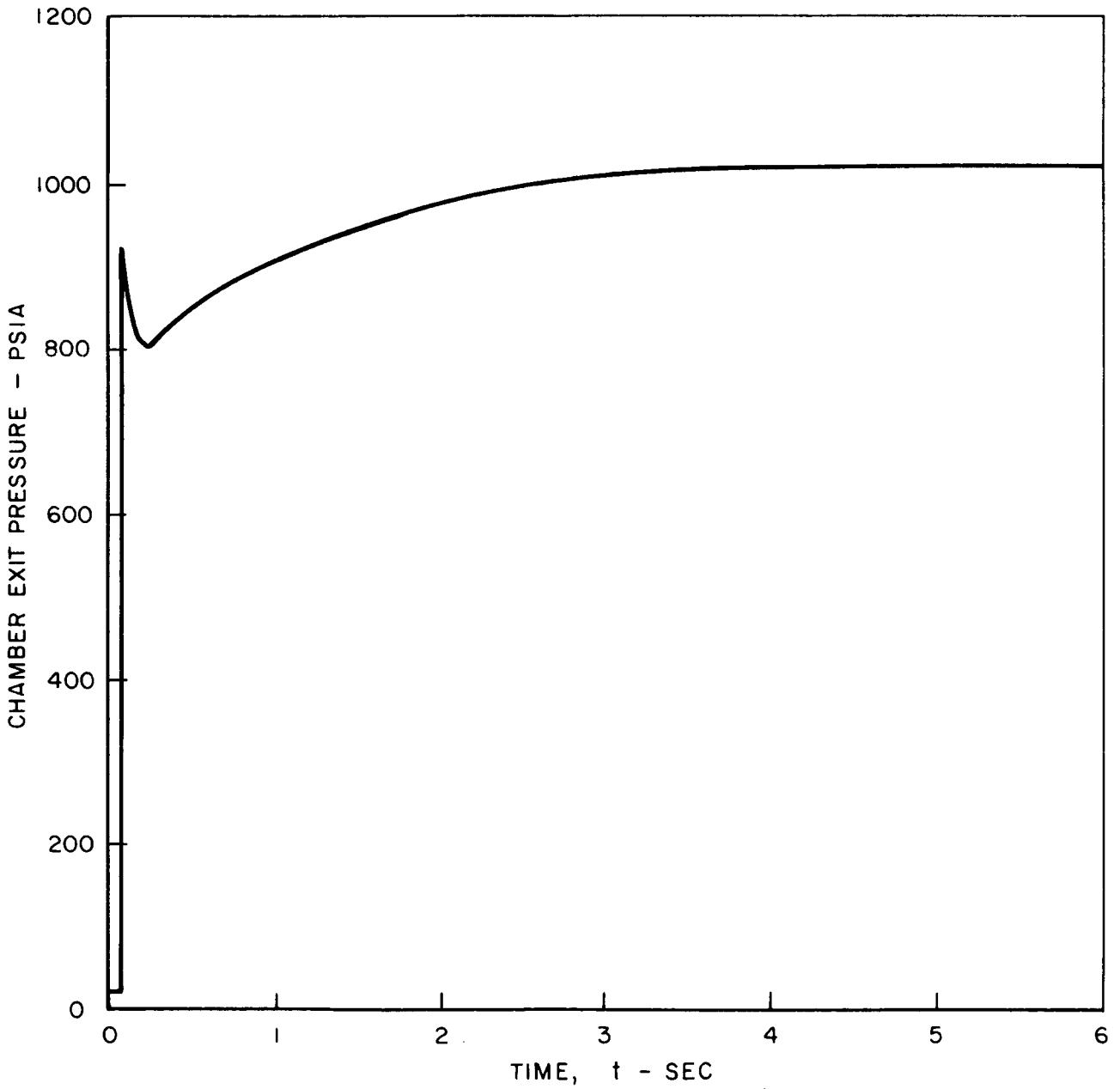
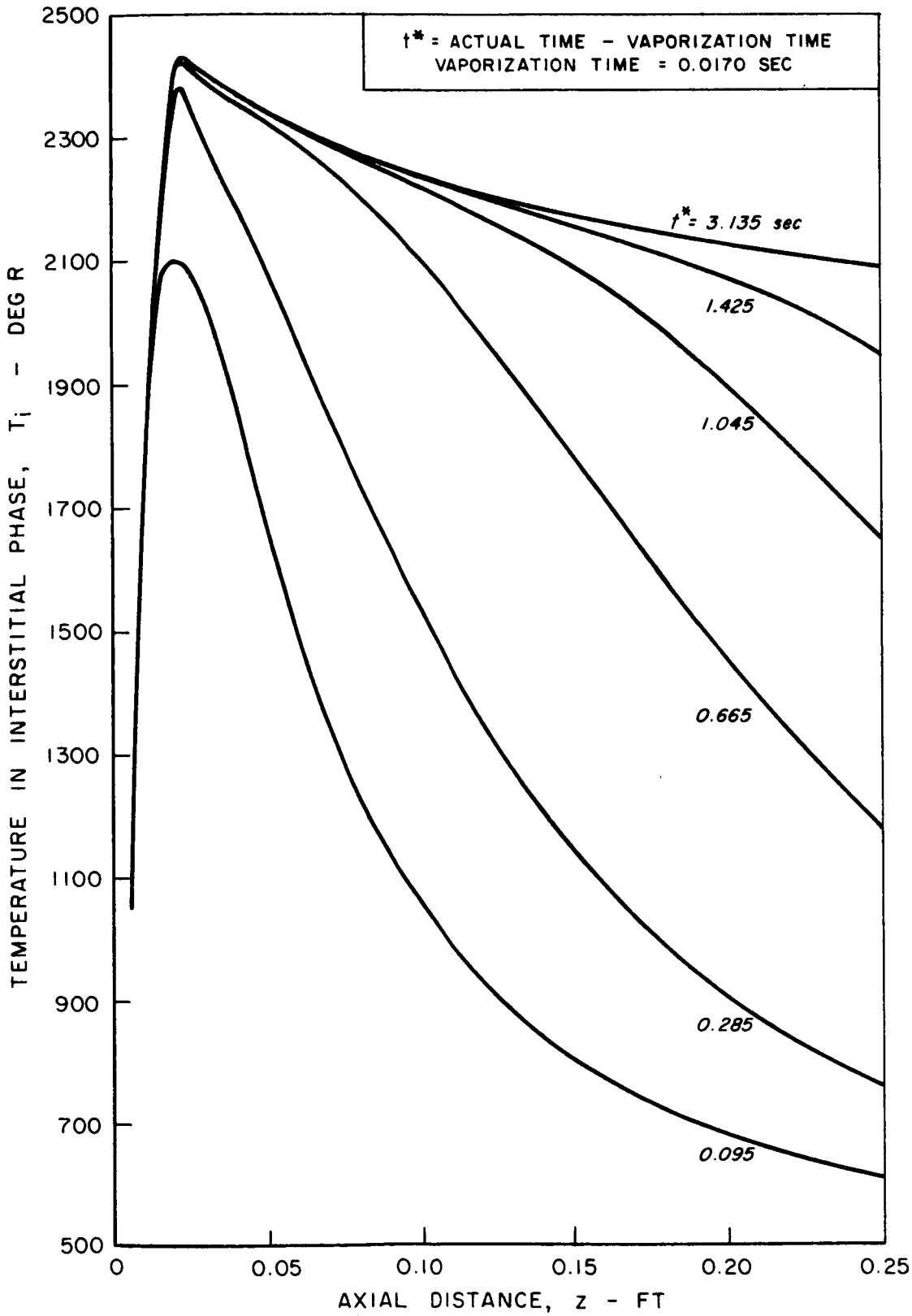


FIG. 36

# TRANSIENT AXIAL TEMPERATURE PROFILES

P (FEED PRESSURE) = 974 PSIA
G = 6.29 LB/FT$^2$- SEC
STANDARD BED CONFIGURATION (SEE TEXT)



# VARIATION OF TEMPERATURE WITH TIME AT SEVERAL AXIAL POSITIONS

P(FEED PRESSURE) = 974 PSIA G = 6.29 LB/FT$^2$-SEC STANDARD BED CONFIGURATION (SEE TEXT)

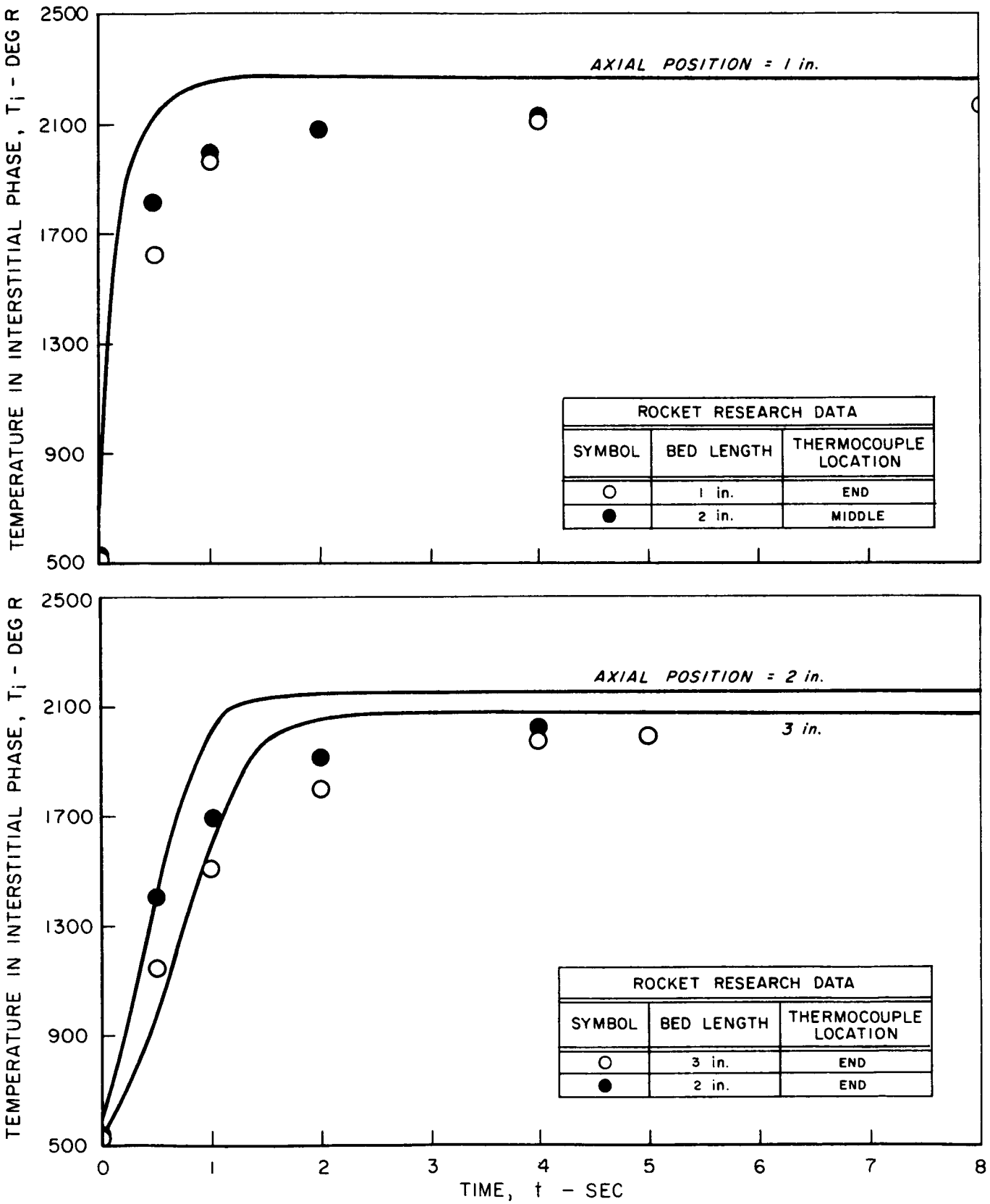
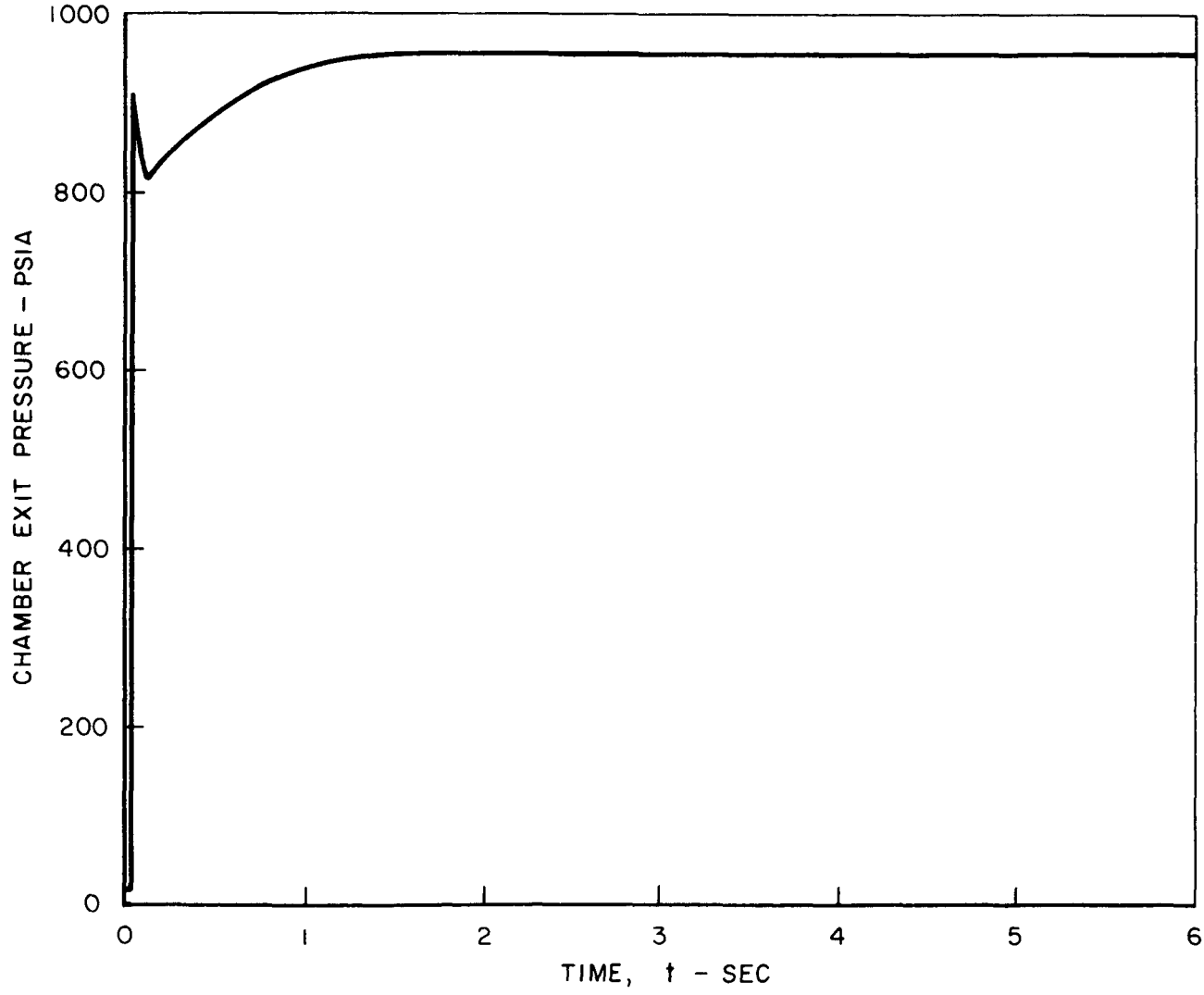


FIG. 38

# VARIATION OF CHAMBER EXIT PRESSURE WITH TIME

P (FEED PRESSURE) = 974 PSIA
G = 6.29 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)



# TRANSIENT AXIAL TEMPERATURE PROFILES

P (FEED PRESSURE) = 217.9 PSIA
G = 1.52 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

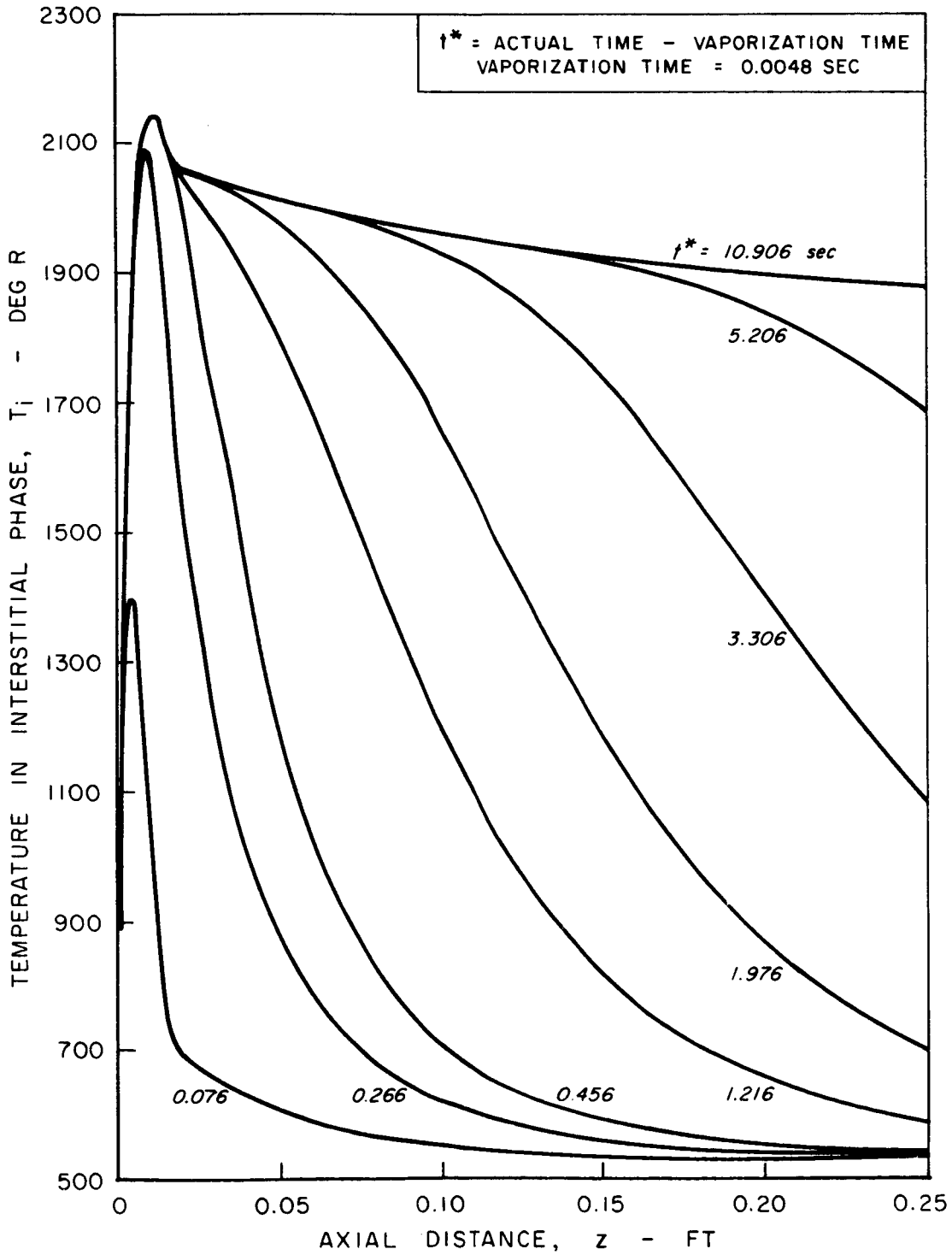
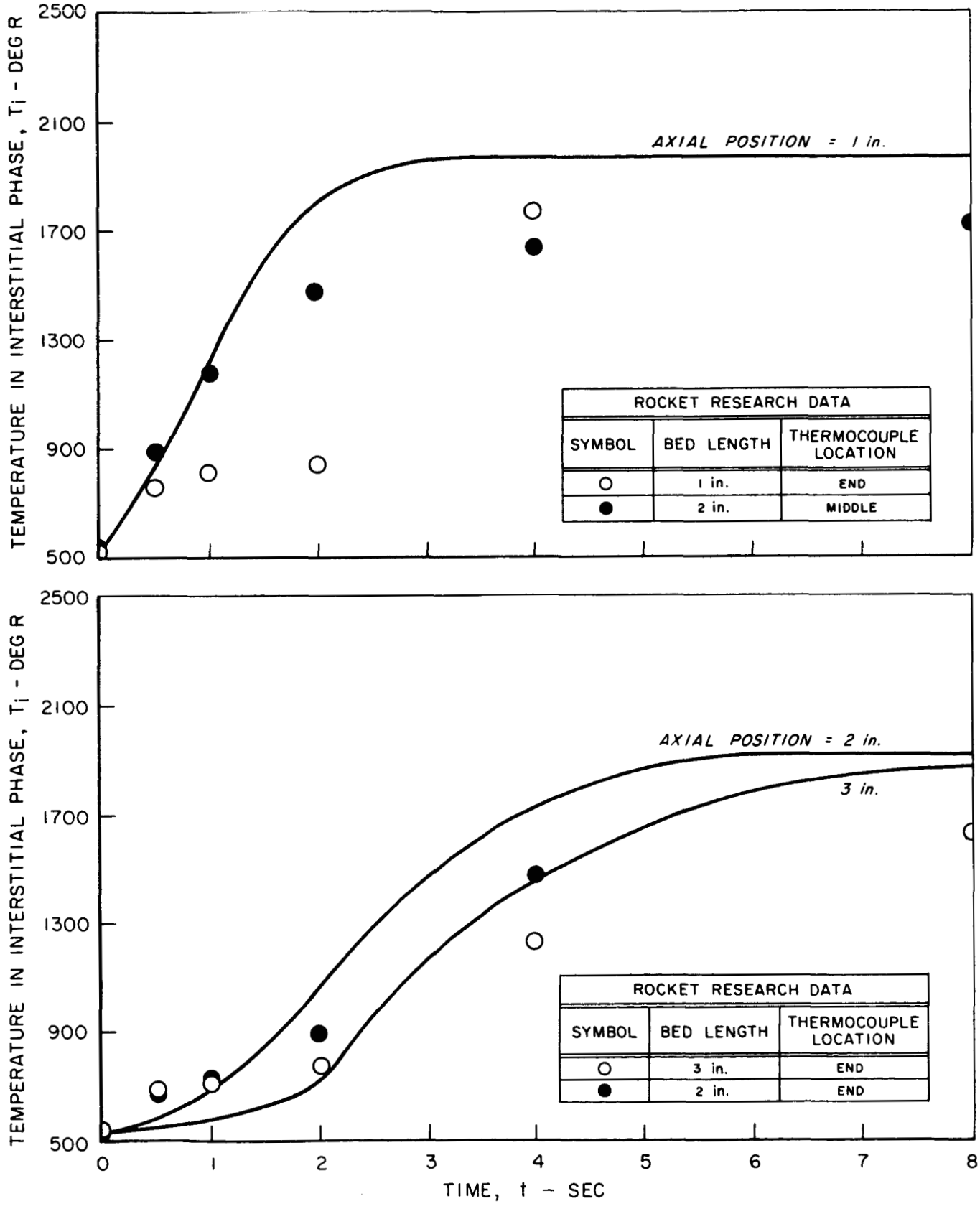


FIG. 40

# VARIATION OF TEMPERATURE WITH TIME AT SEVERAL AXIAL POSITIONS

P (FEED PRESSURE) = 217.9 PSIA  G = 1.52 LB/FT$^2$ - SEC  STANDARD BED CONFIGURATION (SEE TEXT)



# VARIATION OF CHAMBER EXIT PRESSURE WITH TIME

P (FEED PRESSURE) = 217.9 PSIA
G = 1.52 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

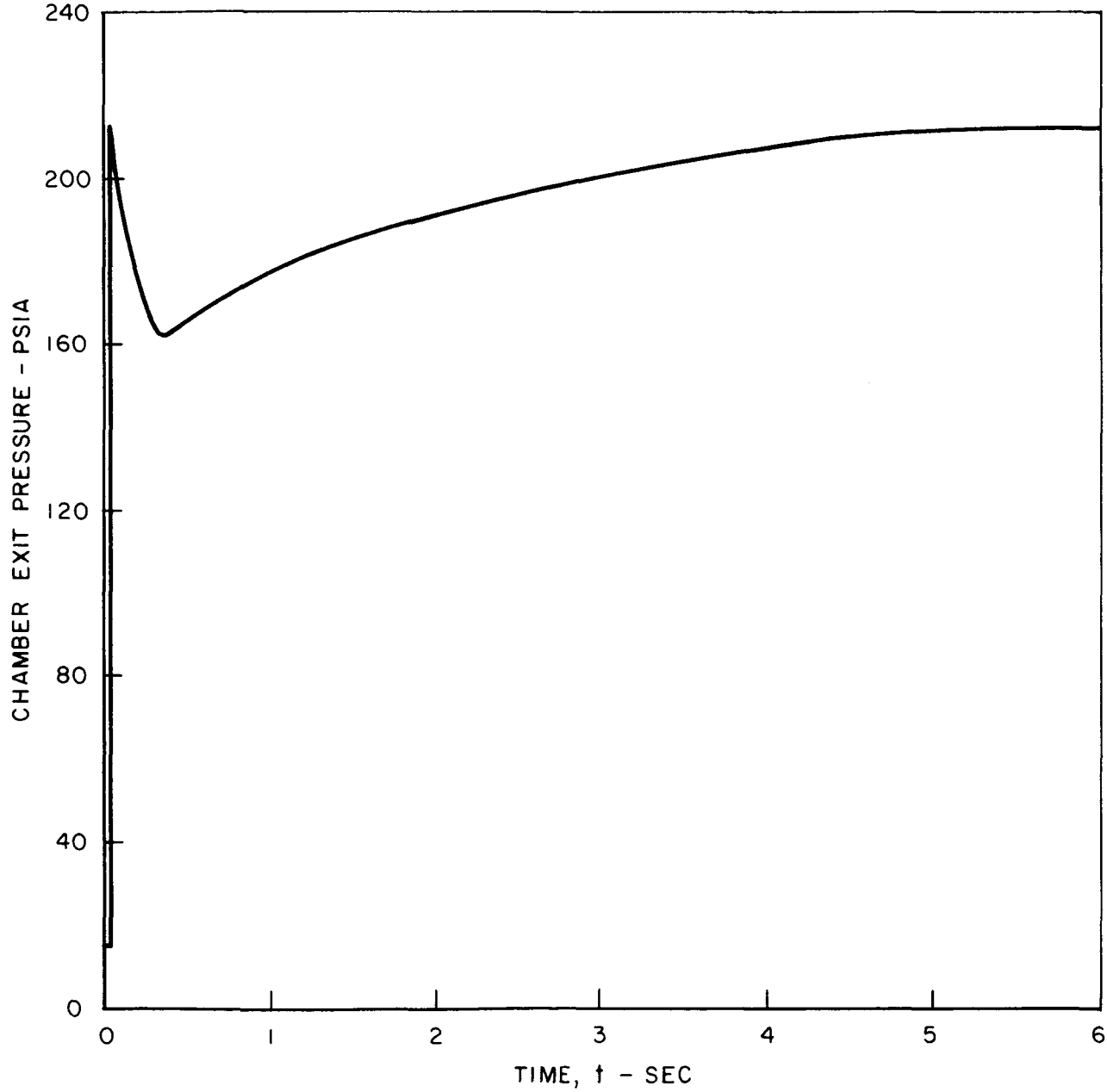


FIG. 42

# TRANSIENT AXIAL TEMPERATURE PROFILES

P(FEED PRESSURE) = 111.4 PSIA
G = 1.51 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

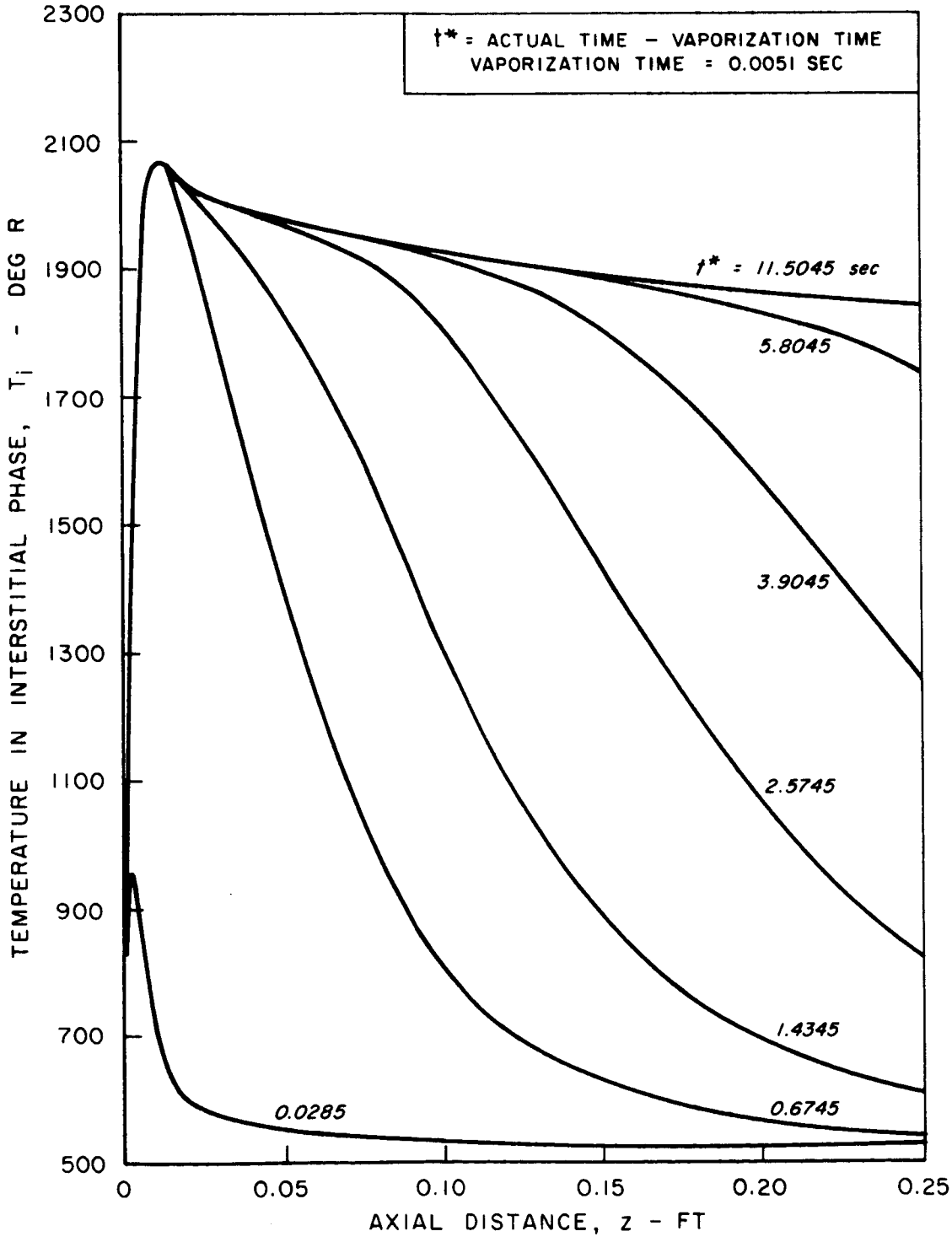


FIG. 43

# VARIATION OF TEMPERATURE WITH TIME AT SEVERAL AXIAL POSITIONS

P(FEED PRESSURE) = 111.4 PSIA  G = 1.51 LB/FT$^2$ - SEC  STANDARD BED CONFIGURATION (SEE TEXT)

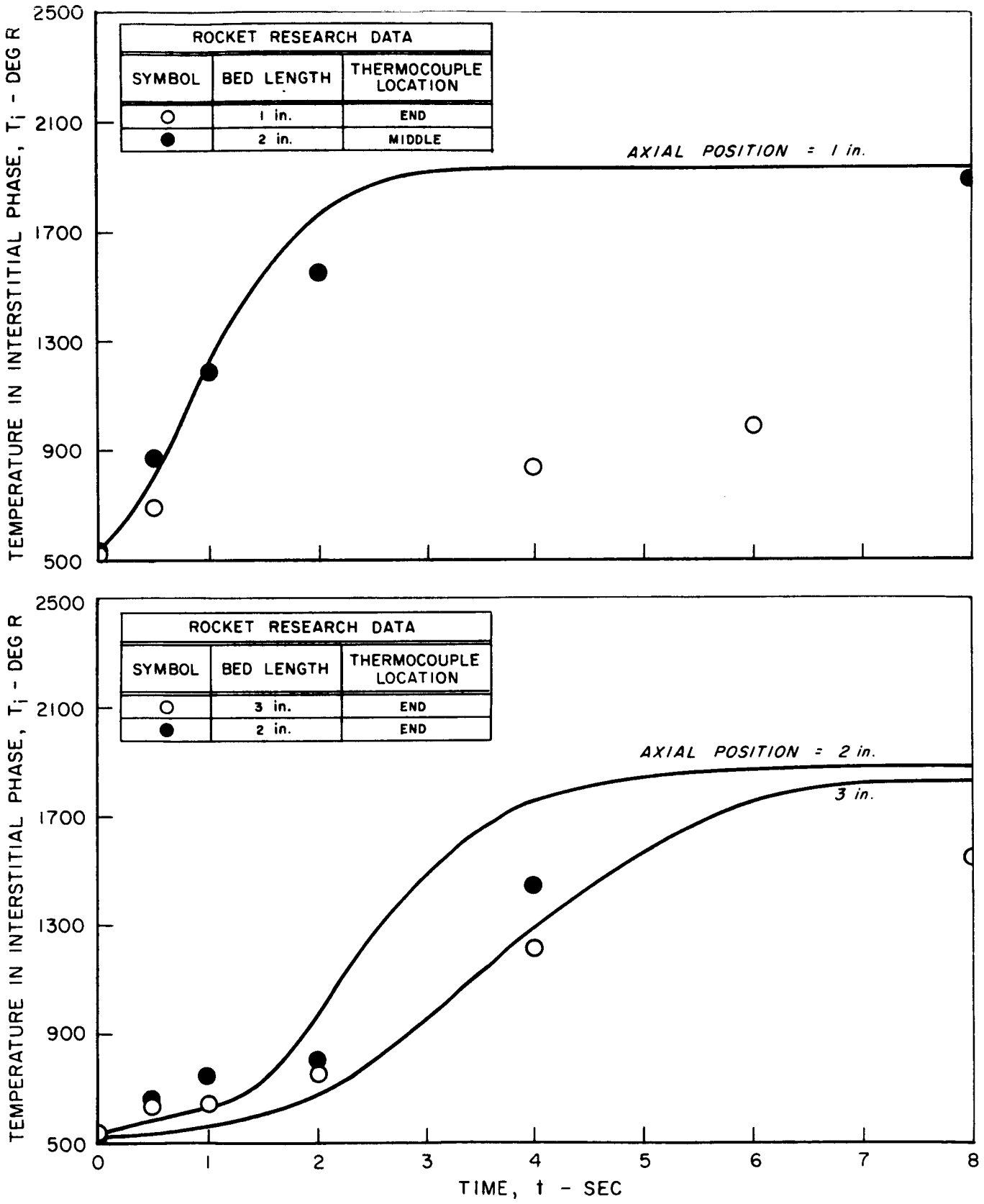


FIG. 44

# VARIATION OF CHAMBER EXIT PRESSURE WITH TIME

P (FEED PRESSURE) = 111.4 PSIA
G = 1.51 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

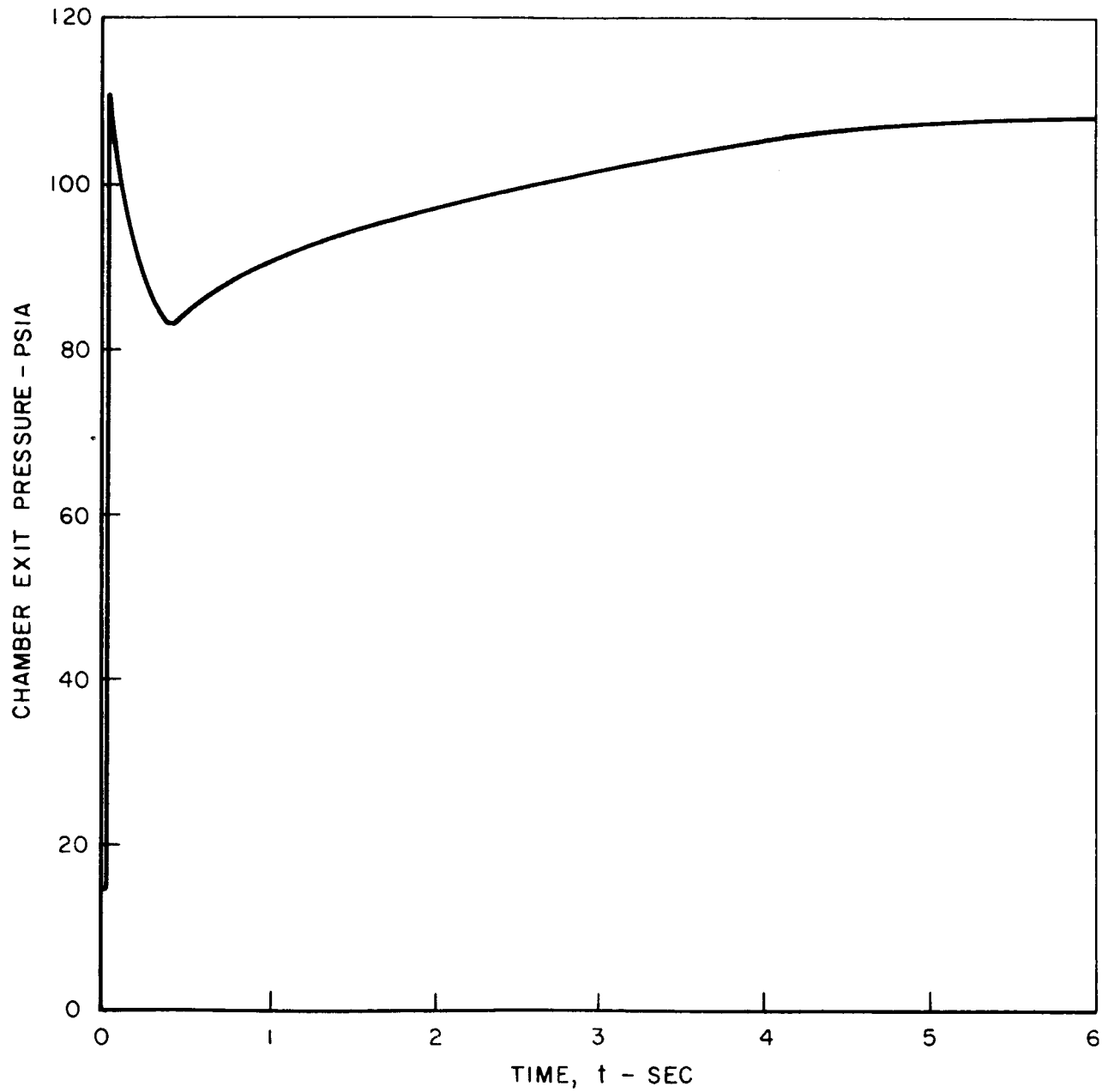


FIG. 45

# VARIATION OF EXIT GAS TEMPERATURE WITH TIME FOR VARIOUS FEED PRESSURES

G = 3.0 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)

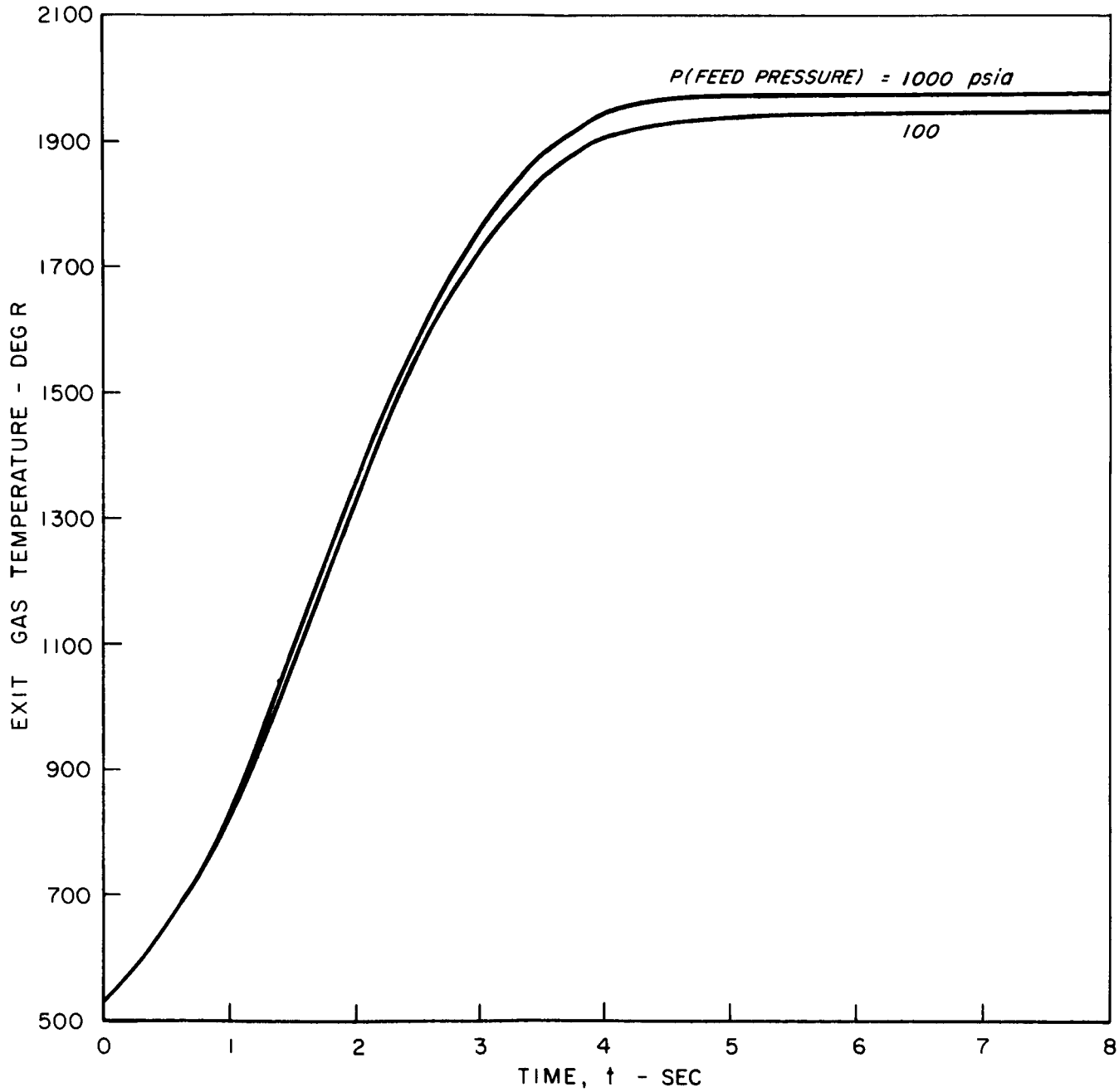


FIG. 46

# VARIATION OF EXIT GAS TEMPERATURE WITH TIME FOR VARIOUS MASS FLOW RATES

P(FEED PRESSURE) = 100 PSIA

STANDARD BED CONFIGURATION (SEE TEXT)

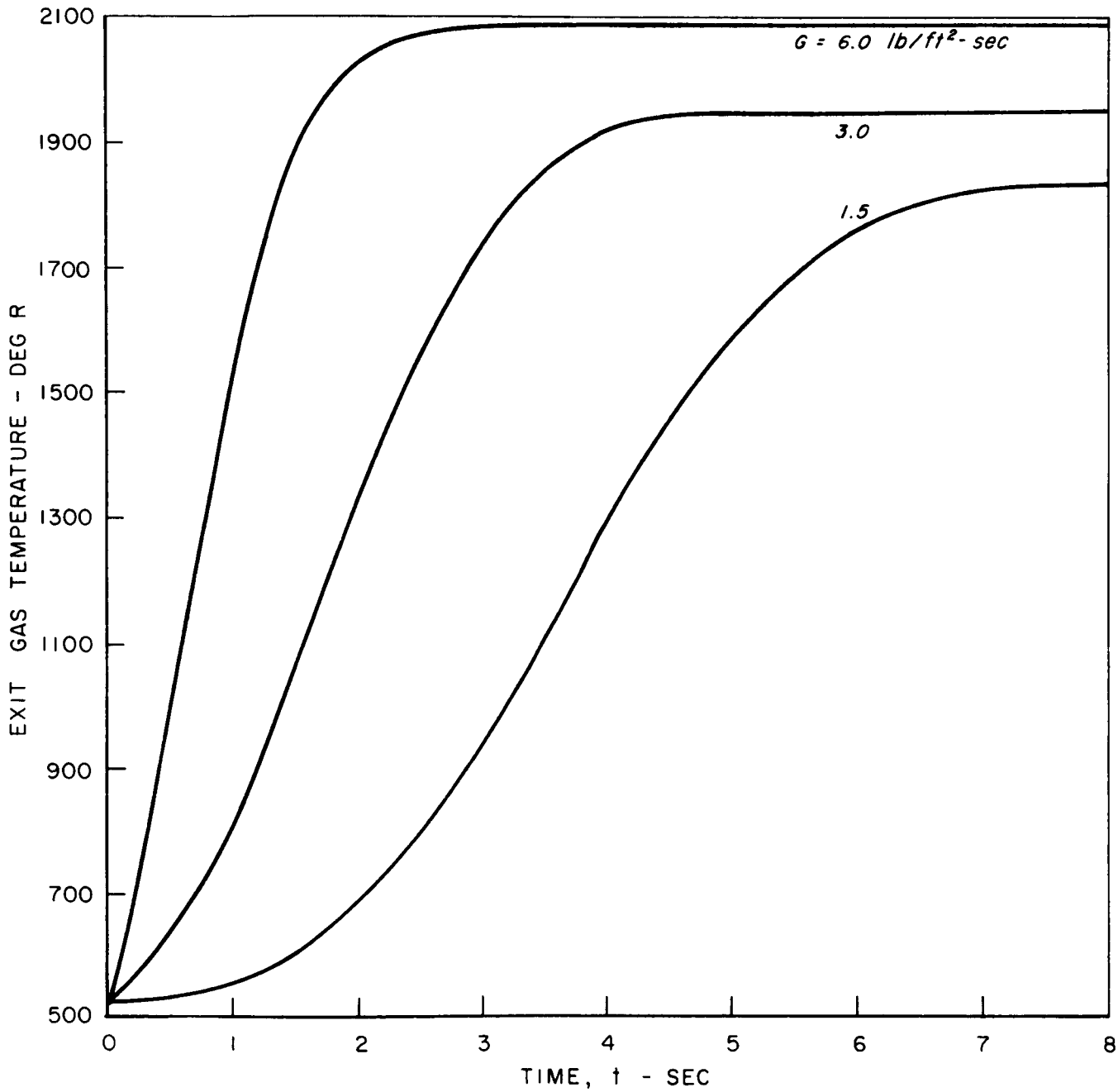
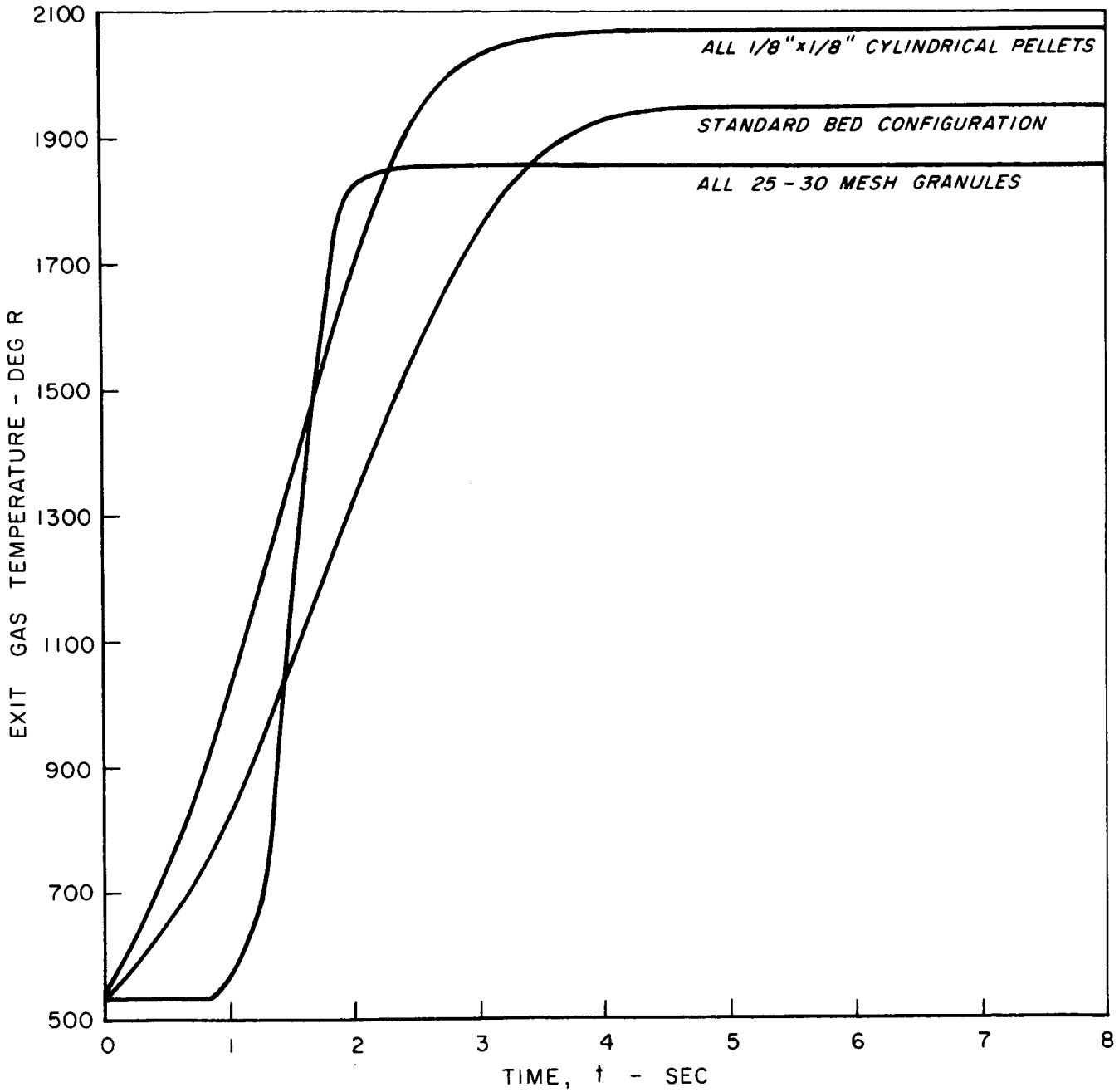


FIG. 47

# VARIATION OF EXIT GAS TEMPERATURE WITH TIME FOR VARIOUS CATALYTIC BED CONFIGURATIONS

P (FEED PRESSURE) = 100 PSIA

G = 3.0 LB/FT$^2$ - SEC



# VARIATION OF EXIT GAS TEMPERATURE WITH TIME FOR VARIOUS HYDRAZINE AXIAL INJECTION PROFILES

P (FEED PRESSURE) = 100 PSIA

BED CONFIGURATION: ALL 25 - 30 MESH GRANULES

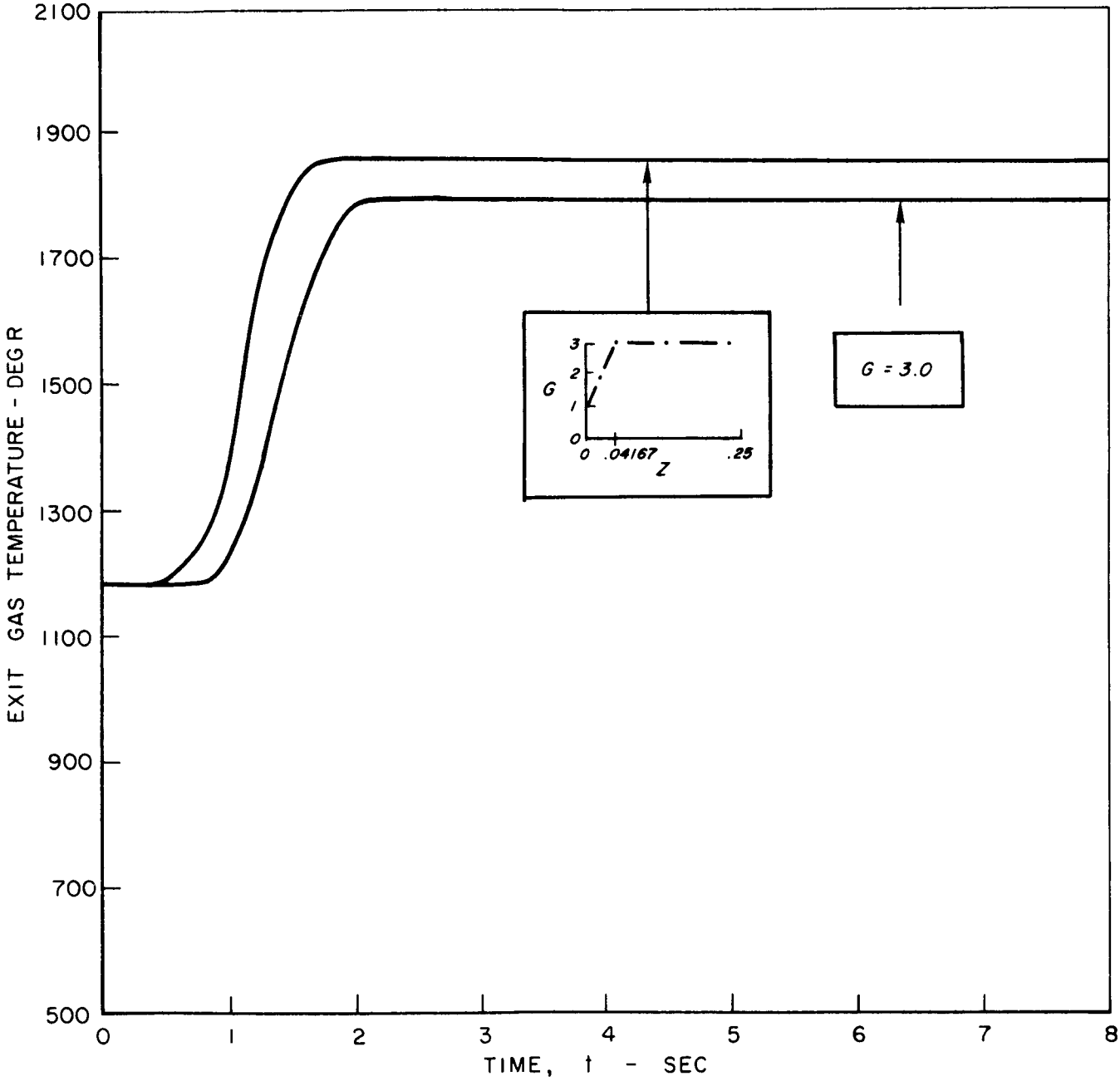
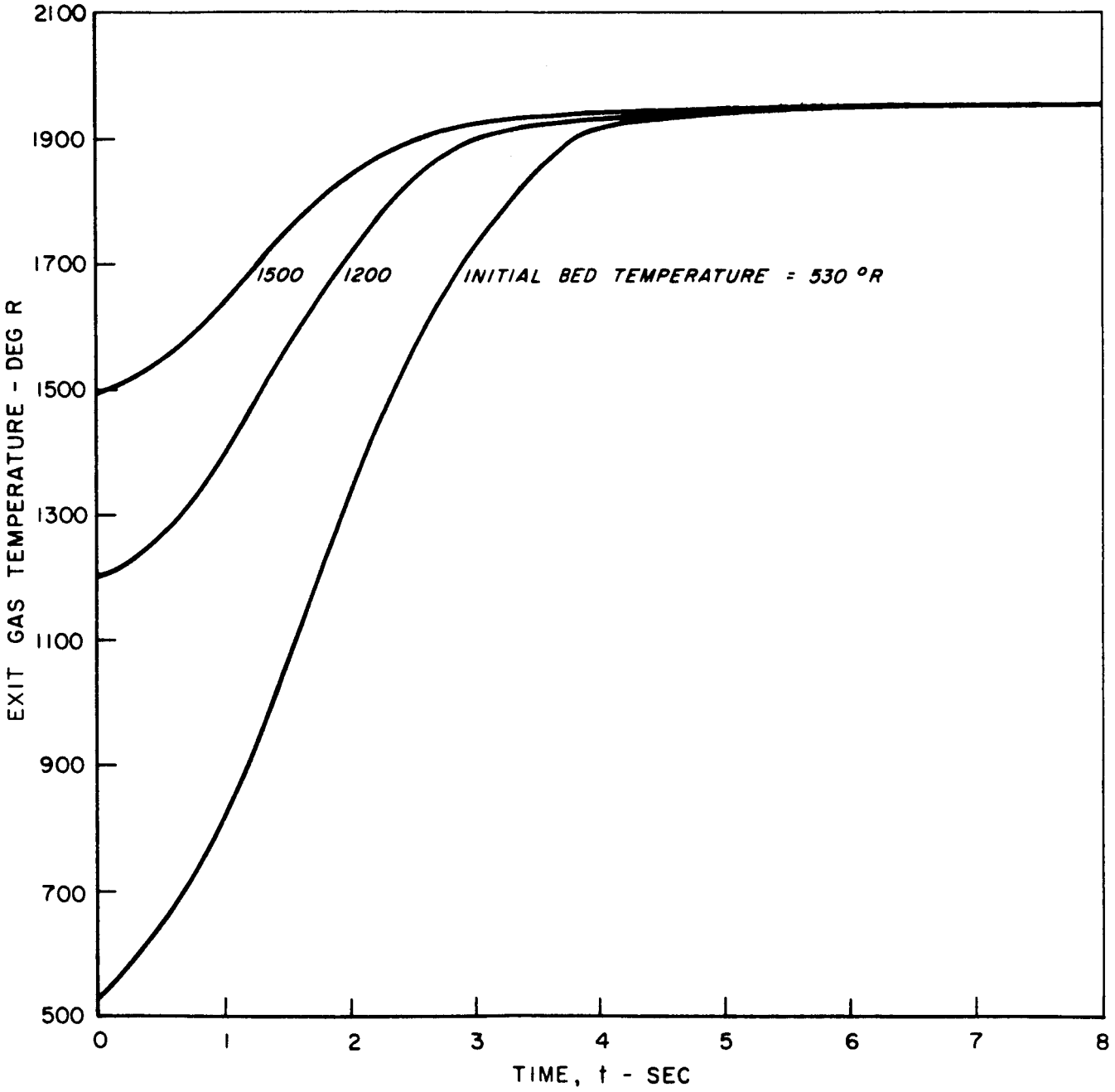


FIG. 49

# VARIATION OF EXIT GAS TEMPERATURE WITH TIME FOR VARIOUS INITIAL BED TEMPERATURES

P(FEED PRESSURE) = 100 PSIA
G = 3.0 LB/FT$^2$ - SEC
STANDARD BED CONFIGURATION (SEE TEXT)



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